U.S. DEPARTMENT OF COMMERCE
National Technical Information Service
PB80-117542

Review of Techniques for Treatment and Disposal of Phosphorus-Laden Chemical Sludges

SCS Engineers, Long Beach, CA

Prepared for

Municipal Environmental Research Lab, Cincinnati, OH Wastewater Research Div

Aug 79

United States
Environmental Protection
Agency

Municipal Environmental Research
Laboratory
Cincinnati OH 45268

EPA-600/2-79-083
August 1979

Research and Development

PB80-117542

EPA

Review of Techniques for Treatment and Disposal of Phosphorus-Laden Chemical Sludges

REPRODUCED BY
NATIONAL TECHNICAL
INFORMATION SERVICE
U.S. DEPARTMENT OF COMMERCE
SPRINGFIELD, VA. 22161

RESEARCH REPORTING SERIES

Research reports of the Office of Research and Development, U.S. Environmental Protection Agency, have been grouped into nine series. These nine broad categories were established to facilitate further development and application of environmental technology. Elimination of traditional grouping was consciously planned to foster technology transfer and a maximum interface in related fields. The nine series are:

1. Environmental Health Effects Research
2. Environmental Protection Technology
3. Ecological Research
4. Environmental Monitoring
5. Socioeconomic Environmental Studies
6. Scientific and Technical Assessment Reports (STAR)
7. Interagency Energy-Environment Research and Development
8. "Special" Reports
9. Miscellaneous Reports

This report has been assigned to the ENVIRONMENTAL PROTECTION TECHNOLOGY series. This series describes research performed to develop and demonstrate instrumentation, equipment, and methodology to repair or prevent environmental degradation from point and non-point sources of pollution. This work provides the new or improved technology required for the control and treatment of pollution sources to meet environmental quality standards.

This document is available to the public through the National Technical Information Service, Springfield, Virginia 22161.

TECHNICAL REPORT DATA

(Please read Instructions on the reverse before completing)

1. REPORT NO. EPA-600/2-79-083	2.	3. RECIPIENT'S ACCESSION NO. PB80-117542
4. TITLE AND SUBTITLE REVIEW OF TECHNIQUES FOR TREATMENT AND DISPOSAL OF PHOSPHORUS-LADEN CHEMICAL SLUDGES		5. REPORT DATE August 1979 (Issuing Date) 6. PERFORMING ORGANIZATION CODE
7. AUTHOR(S) Curtis J. Schmidt, LeAnne E. Hammer, and Michael D. Swayne		8. PERFORMING ORGANIZATION REPORT NO.
9. PERFORMING ORGANIZATION NAME AND ADDRESS SCS Engineers 4014 Long Beach Boulevard Long Beach, California 90807		10. PROGRAM ELEMENT NO. 1BC821(1BC611B), SOS #1,TaskA30) 11. CONTRACT/GRANT NO. 68-03-2432
12. SPONSORING AGENCY NAME AND ADDRESS Municipal Environmental Research Laboratory--Cin.,OH Office of Research and Development U.S. Environmental Protection Agency Cincinnati, Ohio 45268		13. TYPE OF REPORT AND PERIOD COVERED Final 14. SPONSORING AGENCY CODE EPA/600/14

15. SUPPLEMENTARY NOTES

Project Officer: R. V. Villiers (513) 684-7664

16. ABSTRACT

This report summarizes the effects of phosphorus removal by chemical addition on sludge handling and disposal options at full-scale wastewater treatment plants. American and Canadian plants which generate phosphorus-laden chemical sludges were surveyed by questionnaire, and 174 responses were received. Investigations at selected plants that were using a variety of phosphorus removal chemicals, points of chemical addition, and sludge treatment/disposal methods were conducted. The plant operating experiences have shown that all of the various sludge treatment unit processes for thickening, stabilization, conditioning, dewatering, and reduction are adversely affected by phosphorus removal. The adverse effects result from both increases in sludge quantity and changes in sludge characteristics. The adverse impacts are reduced when adequate capacity is available to handle the increased sludge quantity. However, many plants have inadequate capacity, and therefore have been forced to find innovative solutions to problems. This report documents such problem-solving attempts and the results achieved. It also compares the various sludge handling alternatives, and finds that relatively few problems have been encountered with pressure filtration of iron sludges, flotation thickening of iron and aluminum sludges, thermal conditioning of iron sludges, and land disposal of lime sludges. The report contains a bibliography of literature dealing with phosphorus-laden sludges, with an indication as to the scope of each reference.

17. KEY WORDS AND DOCUMENT ANALYSIS

a. DESCRIPTORS	b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group
Chemical removal (sewage treatment) Sludge digestion Sludge drying Sludge disposal	Phosphorus-laden chemical sludges Phosphorus removal (sewage treatment) Sludge treatment and disposal	13B

18. DISTRIBUTION STATEMENT RELEASE TO PUBLIC	19. SECURITY CLASS *(This Report)* UNCLASSIFIED	21. NO. OF PAGES 345
	20. SECURITY CLASS *(This page)* UNCLASSIFIED	22. PRICE A15 A01

EPA Form 2220-1 (Rev. 4-77)

☆ U.S. GOVERNMENT PRINTING OFFICE: 1979 -657-060/5460

1

EPA-600/2-79-083
August 1979

REVIEW OF TECHNIQUES FOR TREATMENT AND DISPOSAL OF PHOSPHORUS-LADEN CHEMICAL SLUDGES

by

Curtis J. Schmidt
LeAnne E. Hammer
Michael D. Swayne

SCS ENGINEERS
Long Beach, California 90807

Contract No. 68-03-2432

Project Officer

R. V. Villers
Wastewater Research Division
Municipal Environmental Research Laboratory
Cincinnati, Ohio 45268

MUNICIPAL ENVIRONMENTAL RESEARCH LABORATORY
OFFICE OF RESEARCH AND DEVELOPMENT
U.S. ENVIRONMENTAL PROTECTION AGENCY
CINCINNATI, OHIO 45268

DISCLAIMER

This report has been reviewed by the Municipal Environmental Research Laboratory, U.S. Environmental Protection Agency, and approved for publication. Approval does not signify that the contents necessarily reflect the views and policies of the U.S. Environmental Protection Agency, nor does mention of trade names or commercial products constitute endorsement or recommendation for use.

FOREWORD

The Environmental Protection Agency was created because of increasing public and governmental concern about the dangers of pollution to the health and welfare of the American people. Noxious air, foul water, and spoiled land are tragic testimony to the deterioration of our natural environment. The complexity of that environment and the interplay between its components require a concentrated and integrated attack on the problem.

Research and development constitute that necessary first step in a solution of the problem, and involve a definition of the problem, the measurement of its impact, and a search for solutions. The Municipal Environmental Research Laboratory develops new and improved technology and systems for the prevention, treatment, and management of wastewater and solid and hazardous waste pollutant discharges from municipal and community sources; for the preservation and treatment of public drinking water supplies; and to minimize the adverse economic, social, health, and aesthetic effects of pollution. This publication is one of the products of that research; a most vital communications link between the researcher and the user community.

This report relates the actual operating experiences of treatment plants which practice phosphorus removal by chemical addition, indicating that all of the various sludge treatment unit processes for thickening, stabilization, conditioning, dewatering, and reduction are adversely affected by phosphorus removal. The more promising methods for handling phosphorus-laden chemical sludges were identified as pressure filtration of iron sludges, flotation thickening of iron and aluminum sludges, thermal conditioning of iron sludges, and land disposal of lime sludges.

Francis T. Mayo
Director
Municipal Environmental Research
Laboratory - Cincinnati

ABSTRACT

Removal of phosphorus from wastewater normally entails precipitation of phosphates by the addition of a chemical, generally either lime or a salt of iron or aluminum. A consequence of phosphorus removal, therefore, is the production of sludge which is laden with chemical precipitates. A survey of 174 municipal plants using chemicals to remove phosphorus from wastewater was conducted in order to quantify the effects of chemical addition on the sludge handling and disposal operations at full-scale plants.

Because of the generation of chemical sludge, phosphorus removal adversely affects a treatment plant in two ways. First, the volume or mass of sludge that must be handled and disposed of is significantly increased. Second, the resulting combined chemical-organic sludges thicken, dewater, and incinerate differently, and often with more difficulty than do organic sludges alone. Both these factors combine to compound the problem of processing and disposing of sludge, and to increase the cost of its handling.

Of the three types of chemicals normally considered for phosphorus removal, iron salts generally appear to have the least overall adverse effect upon subsequent sludge handling. Primary addition of the phosphorus removal chemical often has advantages over secondary addition in terms of the volume and solids concentration of the combined primary/secondary sludge. At existing plants which have gone to phosphorus removal, it has generally not been cost-effective to add a tertiary chemical flocculation and clarification step, unless exceptionally high effluent quality was a goal in addition to phosphorus removal. Neither has it been found economical at existing plants to separate chemical-laden sludges from other sludges for handling if the previous practice has been to combine all the sludges (e.g., primary and secondary, or primary, secondary, and tertiary). Several older plants which have inadequate volume capacity for sludge handling have found that pumping chemical-laden waste activated sludge to the primary clarifier influent for settling with the primary sludge reduces the volume of combined sludge to be treated. At these plants, sludge handling considerations have either been judged to outweigh the problem of deterioration of primary effluent quality, or an increase in secondary clarifier efficiency has counteracted the problem. These conclusions about existing plants should not be

applied to the design of new plants because of the different constraints on costs and options.

Plant operating experiences have shown that all of the various sludge treatment unit processes for thickening, stabilization, conditioning, dewatering, and reduction are adversely affected by phosphorus removal. The adverse impact is reduced when adequate capacity is available to handle the increased sludge quantity. Compared to other alternatives, relatively few problems have been encountered with pressure filtration of iron sludges, flotation thickening of iron and aluminum sludges, thermal conditioning of iron sludges, and land disposal of lime sludges.

This report was submitted in fulfillment of Contract No. 68-03-0268 by SCS Engineers under the sponsorship of the U.S. Environmental Protection Agency. This report covers the period from July 15, 1976, to September 14, 1977. Work was completed as of October 1978.

CONTENTS

Foreword..iii
Abstract..iv
Figures...ix
Tables..xii
Acknowledgments...xviii

1. Introduction...1

2. Project Scope and Methods................................3

3. Phosphorus Removal Impacts on Sludge.....................6

    Types of chemical sludges...............................6
    Quantities of chemical sludges generated................12
    Solids concentration and percent volatile solids of chemical sludges generated...26

4. Prevalence of Various Treatment and Disposal Methods for Chemical Sludges...29

    Introduction..29
    Sludge thickening.......................................29
    Sludge stabilization/reduction..........................29
    Sludge conditioning/stabilization.......................33
    Sludge dewatering.......................................33
    Sludge heat drying......................................33
    Sludge reduction..33
    Sludge final disposal...................................33

5. Thickening of Chemical Sludges...........................34

    Gravity thickening......................................34
    Flotation thickening....................................40

6. Stabilization of Chemical Sludges........................46

    Anaerobic digestion.....................................46
    Aerobic digestion.......................................60
    Composting..67

7. Condition of Chemical Sludges............................71

    Chemical conditioning...................................71
    Thermal conditioning....................................72

Preceding page blank

CONTENTS (continued)

8. Dewatering of Chemical Sludges........................82

Drying beds....................................82
Vacuum filtration..............................88
Drying lagoons.................................93
Pressure filtration............................98
Centrifugation................................105

9. Reduction of Chemical Sludges - Incineration..........112

Introduction..................................112
Questionnaire survey..........................112
Case studies..................................115
Literature....................................116
Conclusions...................................117

10. Disposal of Chemical Sludges - Transport and Land Disposal...119

Introduction..................................119
Questionnaire Survey..........................119
Literature....................................121
Conclusions...................................123

11. State-of-the-Art Appraisal...........................124

Bibliography...133

Appendices

A. Phosphorus-Laden Sludge Management Questionnaire Form..............................141

B. Outline for Collection of Field Data..................149

C. Case Studies..161

Introduction to Case Studies..........................161
Case Study C: South Bend, Indiana.....................161
Case Study D: Sheboygan, Wisconsin....................181
Case Study E: Coldwater, Michigan.....................195
Case Study F: Lakewood, Ohio..........................215
Case Study G: Mentor, Ohio............................242
Case Study H: Brookfield, Wisconsin (Fox River).......256
Case Study I: Midland, Michigan.......................278
Case Study J: Port Huron, Michigan....................293
Case Study K: Pontiac, Michigan.......................308

FIGURES

Number Page

1 Points of chemical addition for phosphorus removal....7

2 Cubic meter (m^3) sludge per kg phosphorus removed (gal sludge per lb removed)....22

3 Cubic meter (m^3) sludge per mil m^3 plant influent....23

4 Kilogram (kg) sludge per kg phosphorus removed....24

5 Kilogram (kg) sludge per m^3 plant influent (lb/MG plant influent)....25

C-1 South Bend, Indiana, wastewater treatment plant flow diagram....166

C-2 Tertiary upflow clarifier configuration, South Bend, Indiana....169

C-3 Flow pattern for South Bend anaerobic digester....173

C-4 Sheboygan, Wisconsin, wastewater treatment plant flow diagram....185

C-5 Sheboygan, Wisconsin, gravity thickener hydraulic and mass balance....186

C-6 Coldwater, Michigan, wastewater treatment plant flow diagram....199

C-7 Hydraulic and solids mass balances for wastewater and sludge treatment operations, Coldwater, Michigan....202

C-8 Flocculation channel, Coldwater, Michigan....206

C-9 Lakewood, Ohio, wastewater treatment plant flow diagram....218

FIGURES (continued)

Number		Page
C-10	Lakewood, Ohio, anaerobic digester configuration	222
C-11	Lakewood, Ohio, hydraulic and solids mass balance during 63 mg/ℓ alum addition (Jan thru Dec 74)	227
C-12	Lakewood, Ohio, hydraulic and solids mass balance before alum addition (Nov 72 thru Oct 73)	228
C-13	1976 anaerobic digestion mass balance during double shift vacuum filter and flash dryer operation, Lakewood, Ohio	234
C-14	Anaerobic digestion mass balance during liquid sludge hauling, Lakewood, Ohio	235
C-15	Greater Mentor wastewater treatment plant wastewater flow diagram	245
C-16	Dual cell gravity thickeners and appurtenances, Mentor, Ohio	252
C-17	Brookfield, Wisconsin, wastewater treatment plant flow diagram	259
C-18	Materials balance for primary wastewater treatment operations at Brookfield, Wisconsin	262
C-19	Materials balance for secondary wastewater treatment operations at Brookfield	263
C-20	Brookfield, Wisconsin pressure filtration and incineration facilities	267
C-21	Pressure filter performance, Brookfield, Wisconsin	268
C-22	Multiple hearth incinerator performance, Brookfield, Wisconsin	270
C-23	Midland, Michigan, wastewater treatment plant flow diagram	281
C-24	Midland, Michigan, sludge treatment flow diagram	286
C-25	Chemical feed points to aeration tanks, Port Huron, Michigan	295

FIGURES (continued)

Number Page

C-26 Port Huron, Michigan, wastewater treatment plant flow diagram....297

C-27 Port Huron, Michigan, gravity thickener hydraulic and mass balance....300

C-28 Wastewater treatment unit process flow diagram, Pontiac, Michigan....315

C-29 Sludge handling unit process flow diagram, Pontiac, Michigan....316

C-30 Pontiac, Michigan, hydraulic and materials balance....318

TABLES

Number Page

1 Results of Plant Survey....................................4

2 Prevalence of Phosphorus Removal Methods (Chemicals and Points of Addition) among Plants Responding to Questionnaire Survey............ 9

3 Combination of Chemical-Laden and Other Sludges for Processing as Practiced by Plants in Questionnaire Survey...11

4 Influence of Plant Size on Type(s) of Chemical(s) Used for Phosphorus Removal among Plants in Questionnaire Survey..................................13

5 SS and BOD Removal Efficiencies and Dry Weights of Suspended Solids Removed at a Hypothetical Activated Sludge Plant...............................15

6 Total Dry Weight of Suspended Solids and Chemical Solids Removed during Treatment Processes at a Hypothetical Activated Sludge Plant................18

7 Theoretical Kilograms of Solids Generated per Kilogram of Phosphorus Removed at a Hypothetical Activated Sludge Plant...............................19

8 Comparison of Theoretical Solids Generation Rates with Results of Questionnaire Survey................26

9 Solids Characteristics of Chemical Sludges with and without Combination with Other Plant Sludges.......................................28

10 Prevalence of Treatment and Disposal Processes for Chemical Sludges among Plants Responding to Questionnaire Survey..................................30

11 Impacts of Chemical Sludge upon Gravity Thickening Performance as Reported in Questionnaire Response....35

TABLES (continued)

Number | Page
12 Impacts of Chemical Sludges upon Flotation Thickener Performance as Reported in Questionnaire Response....41

13 Impacts of Chemical Sludges upon Anaerobic Digester Performance as Reported in Questionnaire Response....47

14 Impacts of Chemical Sludges upon Aerobic Digester Performance as Reported in Questionnaire Response....61

15 Projected Costs of Various Sludge Handling Alternatives Following Aerobic Digestion at Portage Lake, Michigan ($)....65

16 Average Sidestream Characteristics at Grand Haven, Michigan....74

17 Characteristics of Plants in Study of Thermal Conditioning of Chemical Sludges....76

18 Characteristics of Streams and Sidestreams Associated with Sludge Treatment Operations....78

19 Results of Pilot Scale Centrifugation of Midland Thermally Conditioned Iron Sludge....80

20 Comparison of Sidestreams from Plant and Pilot Dewatering Operations....81

21 Impacts of Chemical Sludges upon Drying Bed Performance as Reported in Questionnaire Response....83

22 Polymer Application to Drying Beds....87

23 Changes in Vacuum Filter Performance Reported as a Result of Phosphorus Removal Chemical Sludge Addition....89

24 Impacts of Chemical Sludges upon Drying Lagoon Performance as Reported in Questionnaire Response....95

25 Sludge Treatment/Disposal Methods Used by Plants Practicing Pressure Filtration....100

TABLES (continued)

Number Page

26 Impacts of Chemical Sludges upon Pressure Filter Performance as Reported in Questionnaire Response....101

27 Pilot Filter Press Test at Holland, Michigan....104

28 Centrate Characteristics from Various Scroll Centrifugation Runs....108

29 Impacts of Chemical Sludges on Centrifuge Performance as Reported in Questionnaire Response....110

30 Impacts of Phosphorus-Laden Chemical Sludges on Incinerator Performance as Reported in Questionnaire Response....113

31 Bibliography Information Index....127

32 Key to Bibliography Information Index....131

C-1 Description of Case Study Sites According to Plant Selection Factors....162

C-2 Wastewater Treatment Process Design Parameters, South Bend, Indiana....167

C-3 Summary of 1976 Wastewater Characteristics and Treatment Performance, South Bend, Indiana....168

C-4 Sludge Treatment Process Design Parameters South Bend, Indiana....171

C-5 Chemical Sludge Production and Gravity Thickening, South Bend, Indiana....174

C-6 Anaerobic Digester Gas Production, South Bend, Indiana....176

C-7 A sample of Historical Data Indicating Average Plant Influent Characteristics, Suspended Solids, and BOD Removal, Sheboygan, Wisconsin....182

C-8 Vacuum Filtration and Incineration Performance before and after Phosphorus Removal, Sheboygan, Wisconsin....189

TABLES (continued)

Number		Page
C-9	Influent and Effluent Wastewater Characteristics and Removal Efficiencies, Coldwater, Michigan	196
C-10	General Plant Description Summary, Coldwater, Michigan	200
C-11	Sludge Pumped from Primary Clarifiers to Primary Digester, Coldwater, Michigan	208
C-12	Sludge Transferred from Digesters to Drying Beds (or Lagoon), Coldwater, Michigan	211
C-13	Plant Influent Wastewater Flow Rates, Lakewood, Ohio	216
C-14	Plant SS, BOD, and Phosphorus Concentrations before and during Alum Addition for Phosphorus Removal, Lakewood, Ohio	226
C-15	Plant SS, BOD, and Phosphorus Removals before and during Alum Addition for Phosphorus Removal, Lakewood, Ohio	229
C-16	1976 Sludge Treatment Data during Double Shift Vacuum Filter and Flash Dryer Operation at Lakewood, Ohio	234
C-17	Costs of Treating Alum Sludge during Single and Double Shift Vacuum Filter and Flash Dryer Operation, Lakewood, Ohio	238
C-18	Costs of Alum Sludge Treatment during Single and Double Shift Vacuum Filter and Flash Dryer Operation, Lakewood, Ohio	240
C-19	Influent and Effluent Wastewater Characteristics and Removal Efficiencies, Mentor, Ohio	243
C-20	General Plant Description Summary, Mentor, Ohio	246
C-21	Aerobic Digester Sludge Characteristics, Mentor, Ohio	250
C-22	Performance of Dual Cell Gravity Concentrators, Mentor, Ohio	254

TABLES (continued)

Number Page

C-23 Influent and Effluent Wastewater Characteristics and Removal Efficiencies, Brookfield, Wisconsin.... 257

C-24 General Plant Description Summary, Brookfield, Wisconsin.. 260

C-25 Changes in Primary Clarifier Performance as the Result of Chemical Addition for Phosphorus Removal, Brookfield, Wisconsin.................... 264

C-26 Impacts of Chemical Addition for Phosphorus Removal on Final Clarifiers, Brookfield, Wisconsin.. 266

C-27 Pressure Filter Performance, Brookfield, Wisconsin..273

C-28 Pressure Filter and Incinerator Operational Cost per t (ton) Dry Solids, Brookfield, Wisconsin..277

C-29 Phosphorus Removal Impacts on Plant BOD and SS Removals, Midland, Michigan....................282

C-30 Impacts of Phosphorus Removal: Sludge Conditioning, Thickening, and Dewatering Characteristics -- Variability with Conditioning Method, Thermal Conditioner Temperature, and the Recycling of Tertiary Filter Backwash Water, Midland, Michigan...........287

C-31 A Comparison of the Characteristics of the Sludges Produced with Alum and Ferris Chloride, Midland, Michigan.......................290

C-32 Additional Cost for High Temperature Thermal Conditioning, Midland, Michigan...................292

C-33 Direct Effects of Conditioning Method on Performance and Operational Costs of Solids Handling, Port Huron, Michigan....................304

C-34 Wastewater Characteristics and Removal Efficiencies, Pontiac, Michigan...................310

ACKNOWLEDGMENTS

The project officer for this contract was R. V. Villiers, Ultimate Disposal Section, Wastewater Research Division, Municipal Environmental Research Laboratory, EPA, Cincinnati, OH. SCS Engineers personnel were Lee Hammer, Curt Schmidt, Don Sherman, Mike Swayne, and Mark Montgomery.

We wish to thank the following people for their cooperation and assistance:

Dr. J. B. Farrell
Municipal Environmental Research Laboratory, EPA
Cincinnati, OH

Mr. G. L. Van Fleet and Mr. P. Seto
Sanitary Engineering Branch,
Ontario Ministry of the Environment
Toronto, Ontario

Mr. Charles J. Sluskonis
Greater Mentor Wastewater Treatment Plant
Mentor, OH

Mr. John F. Budde
Fox River Water Pollution Control Center
Brookfield, WI

Mr. J. Michael Jeter
Bureau of Wastewater
South Bend, IN

Mr. John Hennessey and Mr. Jim Spangler
Pontiac Wastewater Treatment Plant
Pontiac, MI

Mr. Larry Marshall
Lakewood Wastewater Treatment Plant
Lakewood, OH

Mr. Willis W. Stubbe
Sheboygan Wastewater Treatment Plant
Sheboygan, WI

Mr. David H. McKay
Coldwater Wastewater Treatment Plant
Coldwater, MI

Mr. Paul S. Hendricks
Port Huron Wastewater Treatment Plant
Port Huron, MI

Mr. Larry Dull and Mr. Arthur Mass
Midland Wastewater Treatment Plant
Midland, MI

TABLES (continued)

Number Page

C-35 General Plant Description Summary, Pontiac, Michigan....312

C-36 Primary Clarifier Waste Stream Characteristics Pontiac, Michigan....317

C-37 Raw and Digested Sludge and Supernatant Characteristics, Pontiac, Michigan....320

C-38 Vacuum Filter Performance, Pontiac, Michigan....322

C-39 Incinerator Performance, Pontiac, Michigan....323

SECTION 1

INTRODUCTION

Starting in the late sixties, regulatory agencies began placing limits upon the phosphorus content of treated wastewater effluents discharged to many surface receiving waters. As a result, over 400 municipal sewage treatment plants in the United States and Canada are now required to implement additional chemical treatment steps to remove 80 percent or more of the phosphorus contained in the raw sewage. There are several chemicals which can be added to precipitate phosphorus in the form of phosphates from: iron salts, aluminum salts, or lime. All such processes generate sludge which is laden with chemical precipitates. This report focuses upon the problems being experienced in municipal sewage sludge management as a result of the addition of these chemical-laden sludges to the sludge treatment processes.

The municipal plants which were studied were operated to achieve at least 80 percent removal of the phosphorus contained in the raw sewage. At higher or lower removal rates, the impacts on sludge which are described in the report would be correspondingly greater or lesser. A few plants which were investigated were not reaching 80 percent removal at the time of the study because of problems in plant design, unusual waste characteristics, plant hydraulic overload, or, in a few cases, inadequate sludge handling capacity.

Included early in this report are sections which describe the changes in sludge volume and characteristics experienced by treatment plants as a result of adding chemicals to remove phosphorus. Emphasis is upon the differences in volume and characteristics experienced by treatment plants as a result of adding chemicals to remove phosphorus. Emphasis is upon the differences in volume and characteristics resulting from use of each of the common phosphorus removal chemicals and the effects of utilizing alternate points of chemical addition in the sewage treatment chain.

Subsequently, the report discusses the effect of the chemical-laden sludges upon typical sludge treatment unit processes (e.g., thickening, digestion, dewatering, incineration, etc.). Problems experienced by various treatment plants are cited, and solutions that were developed are described. The information presented is derived from a questionnaire survey (174 responses) of treatment plants practicing phosphorus removal, field investigation case studies of nine selected plants, and a review of pertinent literature. Whenever possible, treatment plant names and locations are referenced so that readers may obtain additional information directly.

Case study reports of field investigations at selected treatment plants are provided in the appendices. These case studies provide detailed information pertinent to sludge handling at a variety of plants which are generally representative of typical phosphorus removal technology.

It is necessary to warn the reader that sludge generation and management are complex subjects. Every sewage treatment plant is a unique combination of variables in raw sewage characteristics, treatment unit design, operational procedures, etc. For these reasons, it is often difficult to successfully transfer technology or information on sludge volume and mass from one plant to another. However, the experience of others can serve as a background basis for possible solutions to, or avoidance of, similar sludge management problems. It is our hope that this report will be helpful in guiding sewage treatment plant designers and operators toward a realization of potential sludge management problems and solutions when phosphorus removal operations are necessary.

SECTION 2

PROJECT SCOPE AND METHODS

Removal of phosphorus from wastewater normally entails precipitation of phosphates by the addition of a chemical, generally either calcium hydroxide or a salt of iron or aluminum. A consequence of phosphorus removal, therefore, is the production of a sludge which is laden with chemical precipitates. Because of the generation of chemical precipitates, phosphorus removal impacts adversely on a treatment plant in two ways. First, the volume or mass of sludge that must be handled and disposed of is significantly increased. Second, the resulting combined chemical-organic sludges thicken and dewater differently, and often with more difficulty than do organic sludges alone. Both these factors combine to compound the problem of processing and disposing of sludge, and to increase the cost of its handling.

Although increasingly more treatment plants are being required by regulatory agencies to remove phosphorus, the information necessary to competently design or modify engineering works to adequately handle the phosphorus-laden chemical sludges is still largely not available. EPA's Office of Research and Development perceived the need for a project to assemble and evaluate the information needed to detail the most viable methods for handling and disposing of chemical-laden sludges produced by phosphorus removal,and to identify those techniques which are most cost-effective for dealing with them. This no small task because of the variety of types of chemical-laden sludges and the number of methods available for treatment and disposal of them.

It was the objective of this project to conduct extensive data gathering and in-depth evaluation of available hard engineering data, including operating experience of plants practicing phosphorus removal. It is hoped that the information in this report, when combined with other EPA research on the subject will enable the future preparation of satisfactory definitive guidelines for use by federal and state agencies and consulting engineers for treatment and disposal of chemical sludges.

A literature search was conducted to gather all published and unpublished literature pertinent to the management of chemical sludges. Approximately 80 documents were reviewed and found to contain helpful information. These are listed in the bibliography to this report. To make the bibliography more useful, it is preceded by a matrix index which enables the user to determine the subjects covered in each document, and conversely to identify all the documents which include information about a subject of interest.

An extensive questionnaire survey was made in early 1977 of treatment plants practicing phosphorus removal, as identified by state regulatory agencies, EPA regional offices, and the Canadian Ministry of the Environment of Ontario. A copy of the questionnaire mailed out is shown in Appendix A to this report. Questionnaires were mailed to over 400 plants. A number of plants replied that they will have phosphorus removal at some time in the future. A total of 361 plants were identified as currently practicing phosphorus removal. Table 1 shows the number of plants identified as removing phosphorus (361); those that responded to the survey (174) are broken down by state, plus Canada.

TABLE 1. RESULTS OF PLANT SURVEY

State (Name)	Identified Plants in State (No.)	Plants Responding to Survey (No.)
California	2	1
Colorado	1	1
Illinois	22	5
Indiana	26	7
Michigan	91	59
Minnesota	12	4
New York	8	4
Ohio	34	16
Pennsylvania	13	9
Wisconsin	59	26
Texas	1	1
Canada	92	41
TOTAL	361	174

Considering the length of the questionnaire, the response of almost 50 percent is excellent. Phone call follow-ups were required to obtain the response received, and to clarify those questionnaire responses that contained confusing information. The questionnaire asked for a great deal of detailed information, and respondents supplied whatever data they had. Unfortunately, hard engineering and cost data were usually not available to them. Questionnaire responses are tabulated and/or summarized throughout the report where appropriate to the subject being discussed.

Based upon the questionnaire survey, literature information, and recommendations from knowledgeable persons, a total of nine operating treatment plants were selected for field investigations. An average of one person week was spent gathering all available historical and current operating and design data at each selected plant. Emphasis was placed upon determining the effects of the chemical sludge addition on sludge treatment processes, and upon solving the problems encountered. Appendix B shows the field investigation outline form used. Appendices C through K comprise the case studies themselves. Information gathered during the case studies is utilized throughout the report to enhance discussions of various unit treatment processes.

SECTION 3

PHOSPHORUS REMOVAL IMPACTS ON SLUDGE

TYPES OF CHEMICAL SLUDGES

In this report a chemical sludge is any sludge containing chemical precipitates derived from phosphorus removal. The various types of chemical sludges can be defined by the type(s) of chemical(s) used, the type of plant, the point(s) of chemical addition, and whether or not and how the primary, secondary, and tertiary sludges are combined for processing.

It appears that the only chemicals used by full-scale plants for phosphorus removal are lime or salts of iron or aluminum. Some plants use more than one chemical. Lime usually comes to the plant as quicklime (CaO). It is usually slaked to calcium hydroxide lime ($Ca(OH)_2$) before it is added to wastewater. The slaking process requires a great deal of equipment and some additional operator attention and is partly responsible for the relatively infrequent use of lime for phosphorus removal. The iron salts that are used are ferrous sulfate ($FeSO_4$), ferrous chloride ($FeCl_2$), and ferric chloride ($FeCl_3$). Both ferrous sulfate and ferrous chloride are waste products of the pickling process in steel industries. Waste pickle liquor containing ferrous iron is the cheapest source of iron which can be used for phosphorus removal in some areas of the country. Ferric iron can also occasionally be obtained as a by-product from certain industries although this is rare. The aluminum salts that are used for phosphorus removal are aluminum chloride ($AlCl_3$), aluminum sulfate or alum ($Al_2(SO_4)_3$), and sodium aluminate ($NaAlO_2$). Polymers are often used as coagulant aids with any of the phosphorus removal chemicals.

In most modern wastewater treatment plants, several points are available for chemical addition to precipitate phosphorus (see Figure 1). However, differences exist in the points of addition which are available depending on the type of plant and the type of chemical in use.

Primary addition of any of the phosphorus removal chemicals can be practiced. Primary addition of ferrous iron salts is a special case because the phosphates are not always precipitated in the primary clarifier, as is the case with other chemicals.

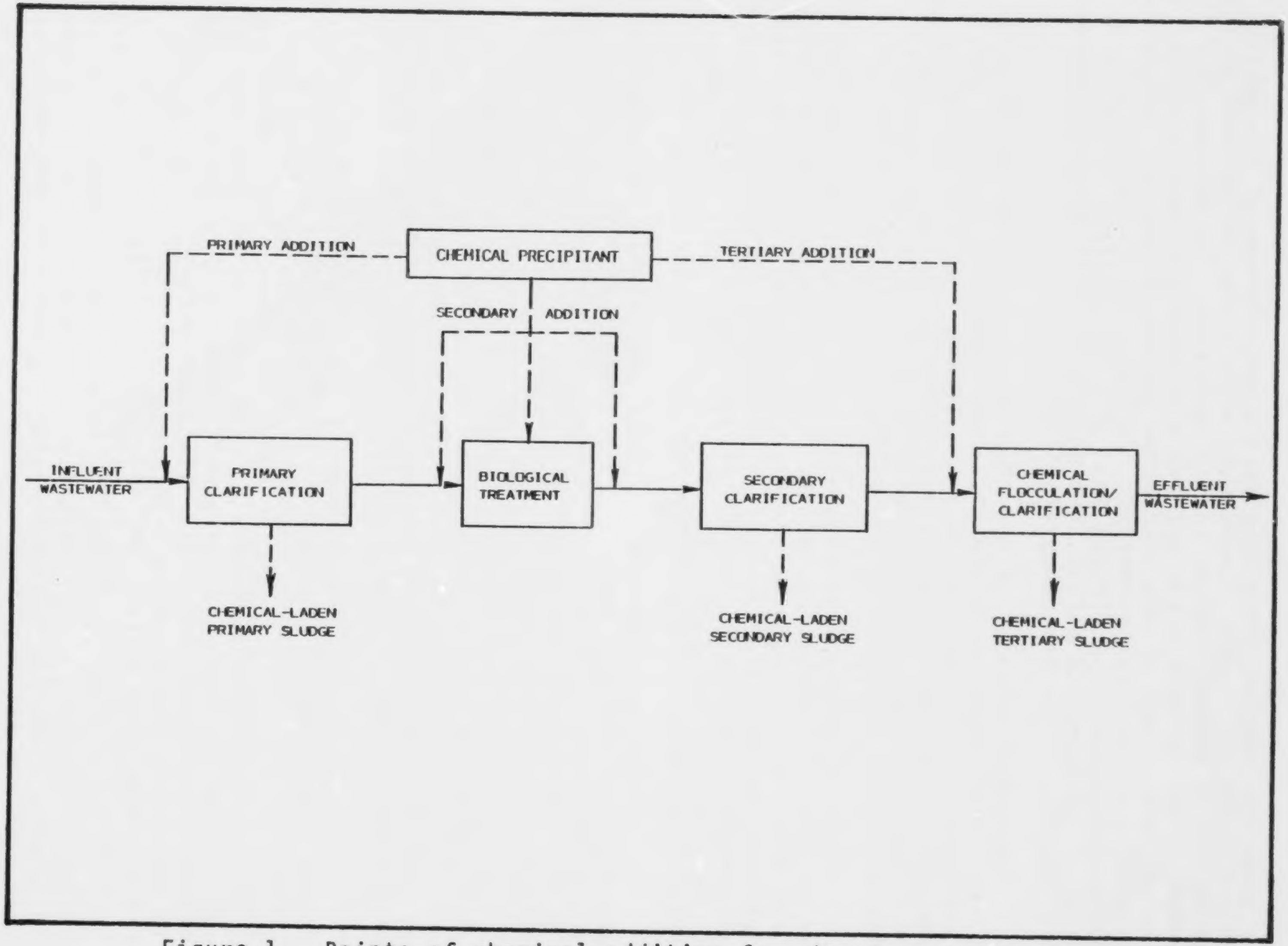

PRIMARY ADDITION
CHEMICAL PRECIPITANT
TERTIARY ADDITION
SECONDARY
ADDITION
INFLUENT
WASTEWATER
PRIMARY
CLARIFICATION
BIOLOGICAL
TREATMENT
SECONDARY
CLARIFICATION
CHEMICAL
FLOCCULATION/
CLARIFICATION
EFFLUENT
WASTEWATER
CHEMICAL-LADEN
PRIMARY SLUDGE
CHEMICAL-LADEN
SECONDARY SLUDGE
CHEMICAL-LADEN
TERTIARY SLUDGE

Figure 1. Points of chemical addition for phosphorus removal.

In order for ferrous iron to affect precipitation of phosphates to a significant extent, the sewage must either be aerated or the pH raised by adding lime or sodium hydroxide. If ferrous iron is added to the primary stage without also adding a base to raise the pH, almost all of the phosphorus passes through the primary treatment without being removed. If the plant has activated sludge treatment, the ferrous ions will be oxidized to ferric ions in the aeration basins, and phosphates will be precipitated and removed in the secondary clarifier.

Secondary addition of any chemical except lime can be practiced. Lime addition to the secondary treatment stage is not practiced because the high pH accompanying lime addition would be detrimental to the biological population of the system. Secondary addition includes addition before, directly into, or just after the activated sludge aeration basins or trickling filters.

Tertiary addition of chemicals in a third stage clarifier is occasionally practiced. Either lime, aluminum salts, or ferrous iron salts are used. Some plants are designed with third-stage treatment primarily because it is effective in removing BOD, SS, and metals in addition to phosphorus. Since it is expensive to build a tertiary clarifier, this practice is uncommon when phosphorus removal alone is the effluent quality objective.

In trickling filter plants, chemicals are seldom added just ahead of, or directly dosed to, the filters themselves, but in activated sludge plants, chemicals are frequently added just prior to or directly into the aeration basins. In trickling filter plants, the chemical would be added before the final clarifiers to avoid staining and clogging of the filters, while in an activated sludge plant it can be advantageous to add the chemicals to the aeration basins to allow utilization of the mixing action which provides dispersion of the chemical in the wastewater. It is sometimes found more effective to add the chemical 2/3 or 3/4 of the way through the aeration basin to minimize floc disruption.

Table 2 presents the results of the questionnaire survey on the prevalence of the various phosphorus removal methods (chemicals and points of addition) among the plants which responded. It also indicates the number of these plants which use a polymer as a flocculation aid. The table shows that chemical addition to the secondary treatment stage is more prevalent than primary or tertiary addition. Sixty-two percent of the plants in the survey practiced secondary chemical addition whereas 26 percent practiced primary addition and six percent added the chemical in a tertiary step involving chemical flocculation and clarification. The remaining six percent used a combination of points of addition. The percentages are basically the same when the plants which did not have either a primary or secondary treatment step are excluded from consideration.

TABLE 2. PREVALENCE OF PHOSPHORUS REMOVAL METHODS (CHEMICALS AND POINTS OF ADDITION) AMONG PLANTS IN QUESTIONNAIRE SURVEY

Point(s) of Chemical Addition	Number of Plants by Plant Type[*]					
	Primary Without Secondary[†]	Aerated Lagoon or Activated Sludge Without Primary[#]	Primary Plus Activated Sludge[**]	Primary Plus Trickling Filter	Total	Total Using Polymer
Primary	10F_3, 3A	NA[‡]	6L, 4F_3,3F_2, 5A, 1L+A	12F_3, 1F_2, 1A	6L, 26F_3,4F_2,9A 1L+A	16F_3, 4A, 1L+A
Secondary	NA	12F_3, 2F_2, 14A	26F_3, 9F_2, 36A, 1F_2+F_3	3F_3, 5A	41F_3, 11F_2, 55A, 1F_2+F3	20F_3, 3F_2, 12A
Tertiary	NA	1L+A	4L, 2F_3, 1A	1F_3, 1A	4L, 3F_3, 2A, 1L+A	3L, 2F_3
Primary and Secondary	NA	NA	1F_2, 1L+F_3, 1F_2+A 2F_3, 1A	1F_3, 1L+A	3F_3, 1F_2, 1A ,1L+F_3, 1L+A, 1F_2+A	2F_3, 1A, 1L+A
Primary and Tertiary	NA	NA	1L+A	NA	1L+A	1L+A
Secondary and Tertiary	NA	1F_3	NA	NA	1F_3	1F_3
Total Number	10F_3, 3A	13F_3, 2F_2, 14A, 1L+A	10L, 34F_3, 13F_2, 43A, 1L+F_3, 2L+A, 1F_2+F_3, 1F_2+A	17F_3, 1F_2, 7A, 1L+A	10L, 74F_3, 16F_2, 67A, 1L+F_3, 4L+A 1F_2+F_3, 1F_2+A	3L, 41F_3, 3F_2, 17A, 3L+A

[*] L = Lime; A = Aluminum salt;
F_3 = Ferric iron salt; F_2 = Ferrous iron salt.

[†] Two plants in this category have aerated lagoons but do not produce secondary sludge.

[#] Three plants in this category have aerated lagoons with sludge removal from final clarifiers; seven plants have the extended aeration modification of the activated sludge process.

[**] Three plants in this category have the extended aeration modification of the activated sludge process

[‡] NA = Not applicable.

Iron salts, either ferric chloride ($FeCl_3$), ferrous chloride ($FeCl_2$), or ferrous sulfate ($FeSO_4$), were the most common chemicals used among the plants in the survey. Fifty-two percent of the plants were using iron. Seventy-five plants used ferric chloride. Both primary and secondary addition of ferric chloride were common, with secondary addition being more prevalent. Tertiary ferric chloride addition was occasionally practiced. One important factor influencing the frequency of iron usage was the availability of waste pickle liquor from steel mills in the Great Lakes Region. All of the 18 plants in the survey which used ferrous iron were obtaining it as an industrial by-product. Most of these plants added the chemical to the secondary stage. The plants which added it to the primary did not accompany it with a base, and therefore the actual phosphate removal occurred mainly in the secondary stage.

Thirty-eight percent of the plants in the survey used aluminum. Secondary addition was much more common than primary, possibly due to difficulty in getting adequate mixing of the aluminum with primary addition. Two plants practiced tertiary addition of aluminum. Fifteen plants reported using lime -- four of them in combination with another chemical. Ten plants added it to the primary stage, while five added it in a tertiary step. With respect to polymer usage, 43 percent of the plants using lime, and 19 percent of the plants using ferrous iron, employed a polymer. Considering the two most commonly used chemicals -- ferric iron and aluminum -- polymer was used in combination with the former 55 percent of the time, but it was used with the latter only 25 percent of the time.

The chemical-laden primary, secondary, or tertiary sludge resulting from lime, aluminum, or iron addition to remove phosphorus can be processed separately from the other sludges produced in the plant, or all of the sludges can be combined for processing (e.g., primary and secondary, or primary, secondary, and tertiary). Sludges can be combined in a sludge-processing unit such as a thickener, digester, holding tank, or mixing tank. Alternatively, primary and secondary sludges are often combined by pumping the waste secondary sludge to the primary clarifier influent where it is mixed with raw sewage. Solids derived from both the secondary sludge and the raw sewage are thus removed together in the primary clarifiers. Table 3 shows that, except in the case of chemical-laden tertiary sludges, it is more common to combine chemical-laden and other sludges than to treat them separately. Tertiary sludges were treated separately at six of the ten plants in the questionnaire survey practicing tertiary chemical addition. At the other four plants, the sludges were combined in or before a digester or a gravity thickener. Among 33 plants practicing primary chemical addition and having both primary and secondary sludges, the chemical-laden primary sludge was combined with the secondary sludge at 31 plants. Thirty-nine percent of these combined by pumping the waste secondary sludge

TABLE 3. COMBINATION OF CHEMICAL-LADEN AND OTHER SLUDGES FOR PROCESSING AS PRACTICED BY PLANTS IN QUESTIONNAIRE SURVEY

Point at Which Chemical-Laden and Other Sludges Were Combined	Number of Plants by Point(s) of Chemical Addition						
	Primary Addition Only	Secondary Addition Only	Tertiary Addition Only	Primary and Secondary Addition	Primary and Tertiary Addition	Secondary and Tertiary Addition	Total
Plants Without Primary or Secondary Clarification							
• No primary sludge	NA*	28	1	NA	NA	1	30
• No secondary sludge	13	NA	NA	NA	NA	NA	13
Plants With Both Primary and Secondary Clarification							
• Sludges not combined	2	10	6	1	0	0	19
• Combined in or before primary clarifier	12	26	0	3	0	0	41
• Combined in or before thickener	13	14	2	1	1	0	31
• Combined in or before digester	6	24	2	3	0	0	35
• Combined in or before dewatering device	0	6	0	0	0	0	6
Total	46	108	11	8	1	1	175†

[*] NA = Not applicable.

[†] One plant with tertiary addition is counted twice because it has no primary sludge, and it has secondary and tertiary sludges which are not combined.

to the primary clarifier influent; 42 percent combined sludges in or before a gravity thickener; and 19 percent in or before a digester. Among 80 plants with secondary chemical addition and both primary and secondary sludges, the sludges were combined at 70 plants. Thirty-seven percent of these plants combined the sludges by pumping to the primary clarifier influent; 20 percent combined sludges in or before a gravity thickener; 34 percent in or before a digester; and 9 percent in or before a dewatering device.

Table 4 examines the influence of plant size on the type(s) of chemical(s) used for phosphorus removal among plants in the questionnaire survey. The table indicates that the majority of plants (74 percent) treats no more than 5 mgd of wastewater. There is no apparent correlation between plant size and chemical used.

QUANTITIES OF CHEMICAL SLUDGES GENERATED

Introduction

The additional sludge generated may be a major concern when implementing chemical addition for phosphorus removal. This section will first present a discussion of stoichiometric relationships to determine the theoretical quantities of chemical precipitates by addition of lime, iron, and aluminum chemicals for phosphorus removal. Second, information taken from literature sources on sludge generation quantities will be presented; and finally, the sludge generation information developed from the treatment plant survey will be presented. Literature and survey information is derived from actual plant experience and will enable comparison of theoretical and actual sludge generation rates. Actual generation is expected to be greater than theoretical rates because of additional solids and BOD capture during chemical treatment.

There are several methods for estimating the mass of sludge which will be generated in a particular plant by phosphorus removal. The best method to use at an existing installation is a plant-scale test. When a full-scale test is infeasible at an existing plant or when an engineer is designing a new plant, a pilot plant test is the next best method. If this is impossible but the actual wastewater to be treated is available, sludge mass should be estimated from jar tests with the wastewater. The jar test results can be corroborated with estimates derived using a calculation method (see Reference 27 for a procedure). The calculation method is most effective when the parameter of chemical dosage can be determined using jar tests. The jar test relates chemical dosage to other parameters; iron or aluminum dosage is related to phosphorus removal; lime dosage is related to alkalinity of the water and the pH level required. If jar tests cannot be carried out, then judgements of these parameters must be made on the basis of past experience.

TABLE 4. INFLUENCE OF PLANT SIZE ON TYPE(S) OF CHEMICAL(S) USED FOR PHOSPHORUS REMOVAL AMONG PLANTS IN QUESTIONNAIRE SURVEY

Chemical(s) Used	Number of Plants by Size					
	0.05-1.0 mgd	1.0-5.0 mgd	5.0-10.0 mgd	10.0-20.0 mgd	over 20.0 mgd	Total
Lime	3	6	0	0	1	10
Ferric iron salt	31	24	7	7	5	74
Ferrous iron salt	3	7	1	3	2	16
Aluminum salt	28	21	9	7	2	67
Lime and Ferric	0	1	0	0	0	1
Lime and Aluminum	0	4	0	0	0	4
Ferrous and Ferric	0	1	0	0	0	1
Ferrous and Aluminum	0	0	1	0	0	1
Total	65	64	18	17	10	174

Estimates of sludge production have been made on a purely empirical basis by noting increases in production at plants which have instituted phosphorus removal. This method is useful only for generalizations because of differences in sewage characteristics and plant design and performance. Lime, iron, or alum addition to an existing plant greatly affects the solids and BOD removal efficiency of the plant; therefore, the increase in sludge production caused by chemical addition depends greatly on what the original removal efficiency was.

Theoretical Sludge Generation Quantities

Knight, Mondoux, and Hambley (40) have calculated sludge generation rates at a hypothetical sewage treatment plant. The hypothetical plant has the following values:

Flow - 3,785 m^3/day (1.0 mgd)

BOD - 200 mg/ℓ

SS - 250 mg/ℓ

Total Dry Weight SS per Day - 947 kg (2086 lb)

P - 10 mg/ℓ

P effluent limitation - 2 mg/ℓ

Table 5 presents, on both dry weight and percentage bases, the wastewater solids removal assumed for this discussion for activated sludge systems: 1) with no chemical addition, 2) with chemical addition for phosphorus removal prior to the primary clarifier, and 3) with tertiary chemical addition for phosphorus removal. The daily weights of dry solids are derived simply by multiplying percentage removals by total suspended solids weights in the influent. Thus, the weights of chemical solids are not included. These calculated weights do not allow for the organics consumed by endogenous respiration or for the soluble BOD converted during aeration into secondary sludge solids. For typical municipal wastewater of the composition shown, respiration and BOD conversion have a negligible effect on the weight values. However, with a high BOD, low suspended solids wastewater, a more detailed calculation would be advisable (see Reference 26 for a procedure).

Lime--

Phosphorus is precipitated by lime as hydroxyapatite, $Ca_5 OH (PO_4)_3$. Assuming 89 percent phosphorus removal and 10 mg/ℓ in the influent, the dry weight of this precipitate from a 3,785 m^3/day (1.0 mgd) plant is 163 kg/day (359 lb/day).

TABLE 5. SS AND BOD REMOVAL EFFICIENCIES AND DRY WEIGHTS OF SUSPENDED SOLIDS REMOVED AT A HYPOTHETICAL ACTIVATED SLUDGE PLANT*

	Primary Treatment			Secondary Treatment			Tertiary Treatment			Total
	Percent Removal		kg/day (lb/day)	Percent Removal		kg/day (lb/day)	Percent Removal		kg/day (lb/day)	kg/day (lb/day)
	SS	BOD	SS Removed	SS	BOD	SS Removed	SS	BOD	SS Removed	SS Removed
No Chemical Addition	50	35	473 (1041)	90	90	378 (833)	—	—	—	851 (1874)
Primary Chemical Addition	75	50	709 (1562)	90	90	142 (312)	—	—	—	851 (1874)
Tertiary Chemical Addition	50	35	473 (1041)	90	90	378 (833)	95	95	47 (104)	898 (1978)

*Weight of chemical solids and additional SS removed during chemical treatment not included.

The percentage of phosphorus removed is essentially a function of pH, with substantial removal occurring as low as pH 9.0. Better than 90 percent removal can be obtained at pH 11.0. The pH required to obtain 80 percent removal of phosphorus is below that at which substantial precipitation of $Mg(OH)_2$ occurs. This affects the thickening and dewatering characteristics of the sludge, since magnesium hydroxide is bulky and dewaters slowly to a very low consistency.

The lime dosage needed to attain any desired pH level depends on wastewater alkalinity, hardness, and the relative quantities of calcium and magnesium ions. In practice, the bulk of the lime added is precipitated in the chemical sludge as calcium carbonate. If all the lime added was precipitated as $CaCO_3$, 1 kg of $Ca(OH)_2$ would yield 1.35 kg of $CaCO_3$. If the lime was used up in magnesium precipitation, 1 kg of $Ca(OH)_2$ would produce 1.35 kg of $CaCO_3$ plus 0.39 kg of $Mg(OH)_2$ for a total of 1.74 kg. One kg of $Ca(OH)_2$ precipitates 1.36 kg of hydroxyapatite.

Although it is not possible, according to Knight, Mondoux, and Hamble (40), to accurately calculate the total amount of solids that will precipitate per pound of lime added, Farrell (26) has presented a procedure for estimating the required lime dose and the quantity of sludge produced for tertiary treatment of a wastewater with lime. All of these authors agree that if the wastewater which will actually be treated is available, the best way to determine lime dose and sludge quantity is by performing jar tests.

For the example of the 3,785 m^3/day (1 mgd) hypothetical plant of Knight, Mondoux, and Hambley (40), assuming that 125 mg/ℓ of $Ca(OH)_2$ is required to increase the pH to 9.5, and that chemical precipitates are produced at the rate of 2.0 kg/kg of $Ca(OH)_2$ added, then the total weight of chemical precipitates for the plant would be 945 kg/day (2082 lb/day). This is roughly equal to the amount of sludge solids produced without chemical addition.

Aluminum Salts--

Phosphorus is precipitated by aluminum ions to form aluminum - $Al_2(SO_4)_3 \cdot H_2O$. A 60 percent excess over the stoichiometric ratio of aluminum to phosphorus is necessary because alum neutralizes alkalinity to lower pH. The excess aluminum is precipitated essentially as aluminum hydroxide ($Al(OH)_3$).

In removing 80 percent of the phosphorus from a 3,785 m^3/day (1 mgd) plant, the alum added will produce 119 kg (262 lb) of $AlPO_4$ and 45 kg (100 lb) of $Al(OH)_3$ for a total weight of chemical solids of 164 kg/day (363 lb/day).

Iron Salts--

Phosphorus is precipitated by ferric ions to form $FePO_4$. The reaction is very similar to aluminum precipitation, with a similar excess of ferric ions needed over the stoichiometric Fe/P ratio. The excess ferric is precipitated as ferric hydroxide, $Fe(OH)_3$. Ferrous iron may also be used. If primary removal is practiced, using ferrous iron and a base to raise the pH, phosphorus precipitates as $Fe_3(PO_4)_2$. However, the stoichiometric ratio is 1.5 for ferrous instead of 1.0 with ferric. If secondary removal is practiced, using ferrous iron in the absence of a base, the plant must have some modification of the activated sludge process. In this situation, the ferrous ion, Fe(II), is converted to the ferric state, Fe(III). Phosphorus precipitates as $FePO_4$ and excess iron precipitates as ferric hydroxide. In removing 80 percent of the phosphorus from a 3,785 m^3/day plant, ferric chloride added in 60 percent excess will produce 147 kg (324 lb) of ferric phosphate and 63 kg (138 lb) of ferric hydroxide for a total weight of chemical solids of 210 kg (462 lb).

Summary--

Adding together the weights calculated for sewage and chemical solids gives the comparison shown in Table 6. The weight of additional SS removed by chemical addition is not included. The table shows the percentage of total plant solids produced at each treatment stage (primary, secondary, and tertiary). For instance, it indicates that the chemical solids produced during lime addition make up over 50 percent of the total plant solids. The dry weight of the total plant solids produced with lime addition is approximately twice that produced with either aluminum or iron. This is true whether the chemical is added at the primary or tertiary stage. The weight of total plant solids produced with either aluminum or iron is approximately equal.

The data presented in Table 6 were used to calculate the theoretical weights of solids generated per kg of phosphorus removed, shown in Table 7. These theoretical quantities will be compared later in this section with values for the weights of sludge per kg of phosphorus removed obtained from the literature and the treatment plant survey. The literature and survey data are based on actual plant operation and include the additional weight of SS removed during chemical treatment.

Sludge Generation Quantities as Reported in the Literature

Recently there have been many studies reported in the literature dealing with the effectiveness of chemical addition for phosphorus removal. Unfortunately, since wastewater effluent quality was the major concern of many of these studies, quantitative determinations of sludge genreation rates were lacking. Therefore the sludge generation values presented here, because they represent only one or two information sources, should not be construed as average sludge generation values. These data should merely be used to augment the sludge generation data from the treatment plant survey found later in this section.

TABLE 6. TOTAL DRY WEIGHT OF SUSPENDED SOLIDS AND CHEMICAL SOLIDS REMOVED DURING TREATMENT PROCESSES AT A HYPOTHETICAL ACTIVATED SLUDGE PLANT*

		No Chemical Addition	Primary Chemical Addition			Tertiary Chemical Addition		
			Lime	Alum	Iron	Lime	Alum	Iron
SS Removed During Primary Treatment	kg/day (lb/day)	473 (1041)	709 (1562)	709 (1562)	709 (1562)	473 (1041)	473 (1041)	473 (1041)
	%[†]	55.5	39.5	69.8	66.8	25.6	44.5	42.7
SS Removed During Secondary Treatment	kg/day (lb/day)	378 (833)	142 (312)	142 (312)	142 (312)	378 (833)	378 (833)	378 (833)
	%	44.5	7.9	14.0	13.4	20.5	35.6	34.1
SS Removed During Tertiary Treatment	kg/day (lb/day)	–	–	–	–	47 (104)	47 (104)	47 (104)
	%	–	–	–	–	2.6	4.4	4.2
Chemical Solids Produced	kg/day (lb/day)	–	945 (2082)	164 (362)	210 (462)	945 (2082)	164 (362)	210 (462)
	%	–	52.6	16.2	19.8	51.3	15.5	19.0
Total Plant Sludge Solids	kg/day (lb/day)	851 (1874)	1796 (3956)	1015 (2236)	1015 (2336)	1843 (4060)	1062 (2340)	1108 (2440)
	%	100	100	100	100	100	100	100

*Weight of additional SS removed during chemical treatment not included.
†Percent of total sludge produced in plant.

TABLE 7. THEORETICAL KILOGRAMS OF SOLIDS GENERATED PER KILOGRAM OF PHOSPHORUS REMOVED AT A HYPOTHETICAL ACTIVATED SLUDGE PLANT*†

Primary Chemical Addition	kg sludge SS/kg P removed
Lime	59
Aluminum	33
Iron	35
Tertiary Chemical Addition	
Lime	61
Aluminum	35
Iron	37

* Derived from Table 6

† Weight of additional SS removed during chemical treatment not included.

Lime--

Primary addition of 200 mg/ℓ of $Ca(OH)_2$ in Ontario, Canada, to achieve an 80 percent reduction in total phosphorus (10.3 to 2.0 mg/ℓ) resulted in an increase in sludge generation from 0.17 kg/m^3 (141 lb/MG) to 0.49 kg/m^3 (4082 lb/MG) (73). The latter generation rate equals 59 kg of sludge per kg of phosphorus removed. Another primary plant mentioned in the same report showed a sludge increase from 0.16 kg/m^3 (1316 lb/MG) to only 0.23 kg/m^3 (1933 lb/Mg) when adding 125 mg/ℓ hydrated lime to achieve 80 percent phosphorus removal. A theoretical discussion of phosphorus removal reports 400 mg/ℓ $Ca(OH)_2$ added to reduce effluent phosphorus from 11.5 to 0.3 mg/ℓ will produce 0.75 kg/m^3 (6290 lb/MG) or approximately 68 kg of sludge per kg of phosphorus removed (19).

Aluminum Salts--

Primary addition of 150 mg/ℓ aluminum sulfate (alum) at the Barrie, Ontario, wastewater treatment plant to achieve 88 percent phosphorus removal resulted in sludge generation of 0.3 kg/m^3 (2428 lb/MG) or 14 kg of sludge generated per kg of phosphorus removed. Secondary alum addition at the Barrie plant of 75 and 100 mg/ℓ generated sludge at 0.39 and 0.31 kg/m^3 (3275 and 2591 lb/MG) respectively. At one Canadian treatment plant, the primary addition of 150 mg/ℓ alum caused a negligible sludge increase from 0.24 to 0.25 kg/m^3 (2033 to 2049 lb/MG); at a second Canadian plant, primary addition of 150 mg/ℓ alum caused a sludge quantity increase of 55 percent from 0.16 to 0.24 kg/m^3 (1316 to 2033 lb/MG) (73).

Iron Salts--

A Canadian activated sludge plant adding 37.5 mg/ℓ ferric chloride to achieve over 95 percent phosphorus removal, experienced an increase in sludge production from 6,200 m^3/mil m^3 to 7,600 m^3/mil m^3 with primary addition of $FeCl_3$ and 7,682 m^3/mil m^3 with secondary addition of $FeCl_3$ (2). Primary addition of 20 mg/ℓ $FeCl_3$ at Sarnia, Ontario, wastewater treatment plant to reduce the effluent phosphorus concentration by 80 percent (5.71 to 1.16 mg/ℓ), resulted in a sludge production of 0.2 kg/m^3 (1,643 lb/MG), or 43 kg of sludge per kg of phosphorus removed (30). The previously mentioned EPA Technology Transfer Publication predicts that an 80 mg/ℓ $FeCl_3$ dosage used to reduce a phosphorus concentration from 11.5 to 0.3 mg/ℓ will produce 0.32 kg/m^3 (2,662 lb/MG) or 28.5 kg of sludge per kg of phosphorus removed (19).

Sludge Generation Quantities as Reported by the Treatment Plant Survey

Approximately 100 plants responding to the questionnaire survey provided information pertinent to sludge volumes and/or weight. Often only fragmentary data or crude estimates were provided. In addition, it is difficult to compare sludge generation by different treatment plants because such individual factors as raw sewage characteristics, plant design, etc., may have a profound effect and distort the results. Nevertheless, the responses were analyzed to look for confirmation of the theoretical and literature information presented in the previous subsections. The questionnaire data are displayed in the form of bar graphs in Figures 2 through 5. The bars indicate average values based on the responses of several treatment plants. Each bar displays a different type of information. Therefore, the value for each bar is generally the average of data from a different set of plants than was the basis for any other bar. This explains some of the inconsistencies in the results.

Figure 2 summarizes the reported sludge volume generated per unit mass of phosphorus removed. The numbers refer to sludge volume before thickening. The following trends are apparent:

- Lime addition generates roughly twice the volume of sludge as does alum or iron addition
- Chemical additions to the secondary treatment process generate substantially more sludge volume than chemical additions to the primary treatment process. However, if the secondary sludge is recycled back to the primary tank, the overall plant sludge volume is not appreciably more than if the chemicals were initially added to the primary treatment process. In other words, there appears to be a definite disadvantage, in terms of total raw sludge volume generated, to handling the secondary sludge separately. Presumably, the secondary sludge thickens in the primary tank.

Figure 3 summarizes the reported volume of sludge generated per unit volume of plant influent. As in Figure 2, the numbers refer to sludge volume before thickening. The trends found in Figure 3 are similar to the Figure 2 results. In addition, sufficient questionnaire data was available to compare increases in sludge volume resulting from the chemical addition with sludge volumes generated prior to the chemical addition. Increases in sludge volume resulting from the addition of iron, aluminum, or lime to the primary clarifiers were:

- 25 percent for iron salt additions
- 58 percent for aluminum salt addition
- huge for lime addition.

As previously stated, the results shown are averages of the values reported by many plants and should not be used for predicting actual sludge volume at a specific plant since many other plant-specific variables affect sludge volume. It should also be noted that the huge lime sludge volumes reported are greatly decreased by sludge thickening.

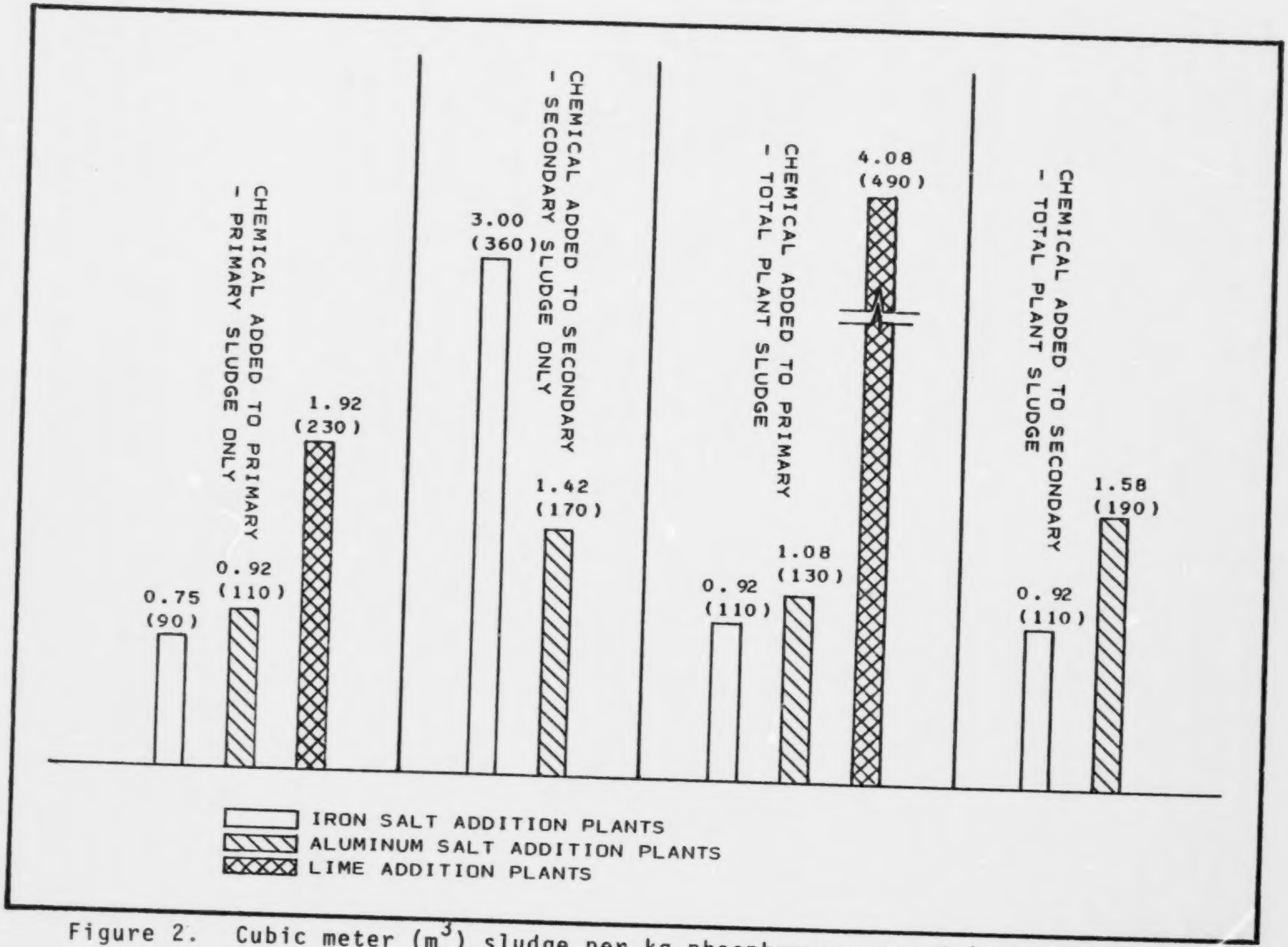


Figure 2. Cubic meter (m^3) sludge per kg phosphorus removed (gal sludge per lb removed).

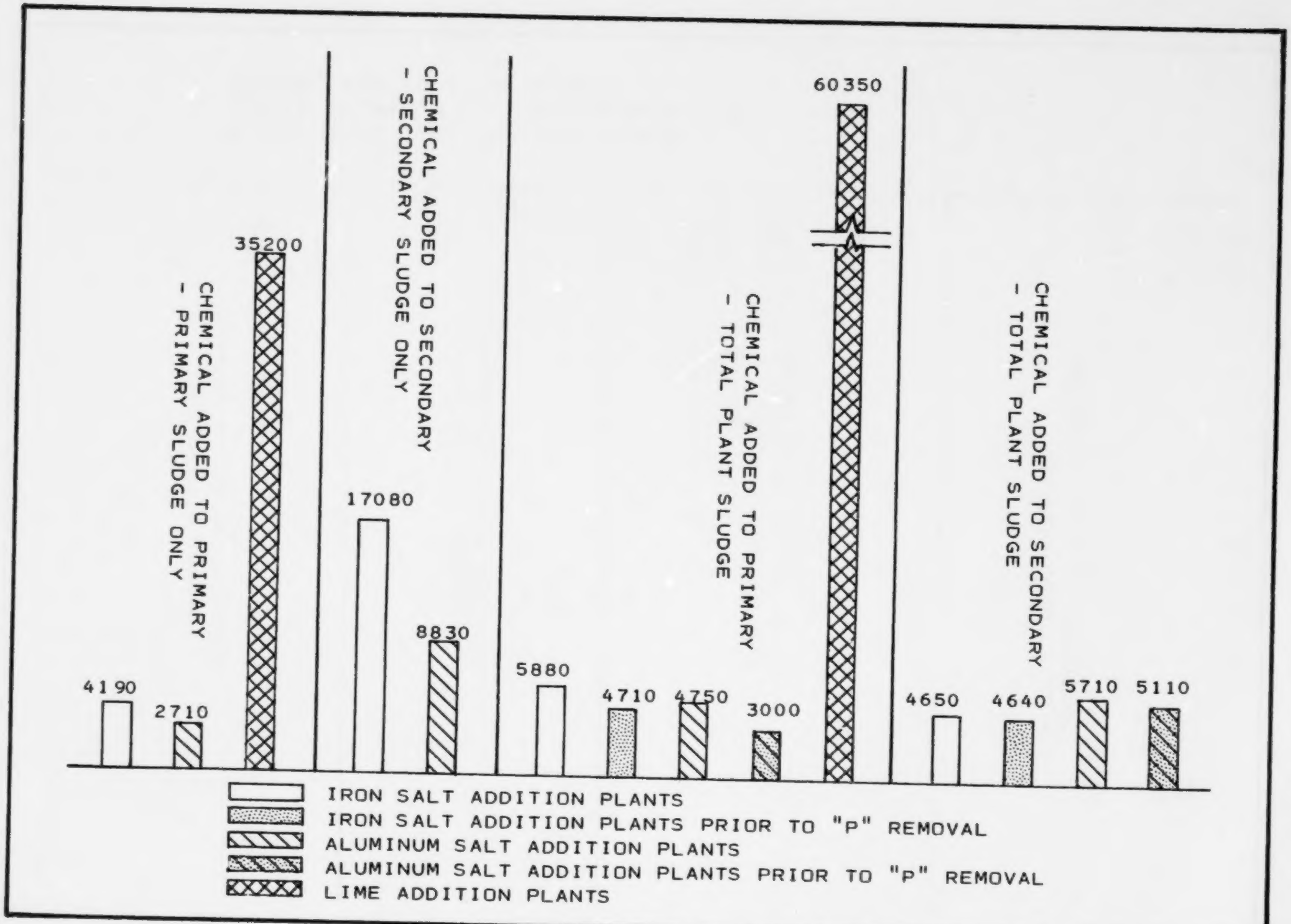


Figure 3. Cubic meter (m3) sludge per mil m^3 plant influent.

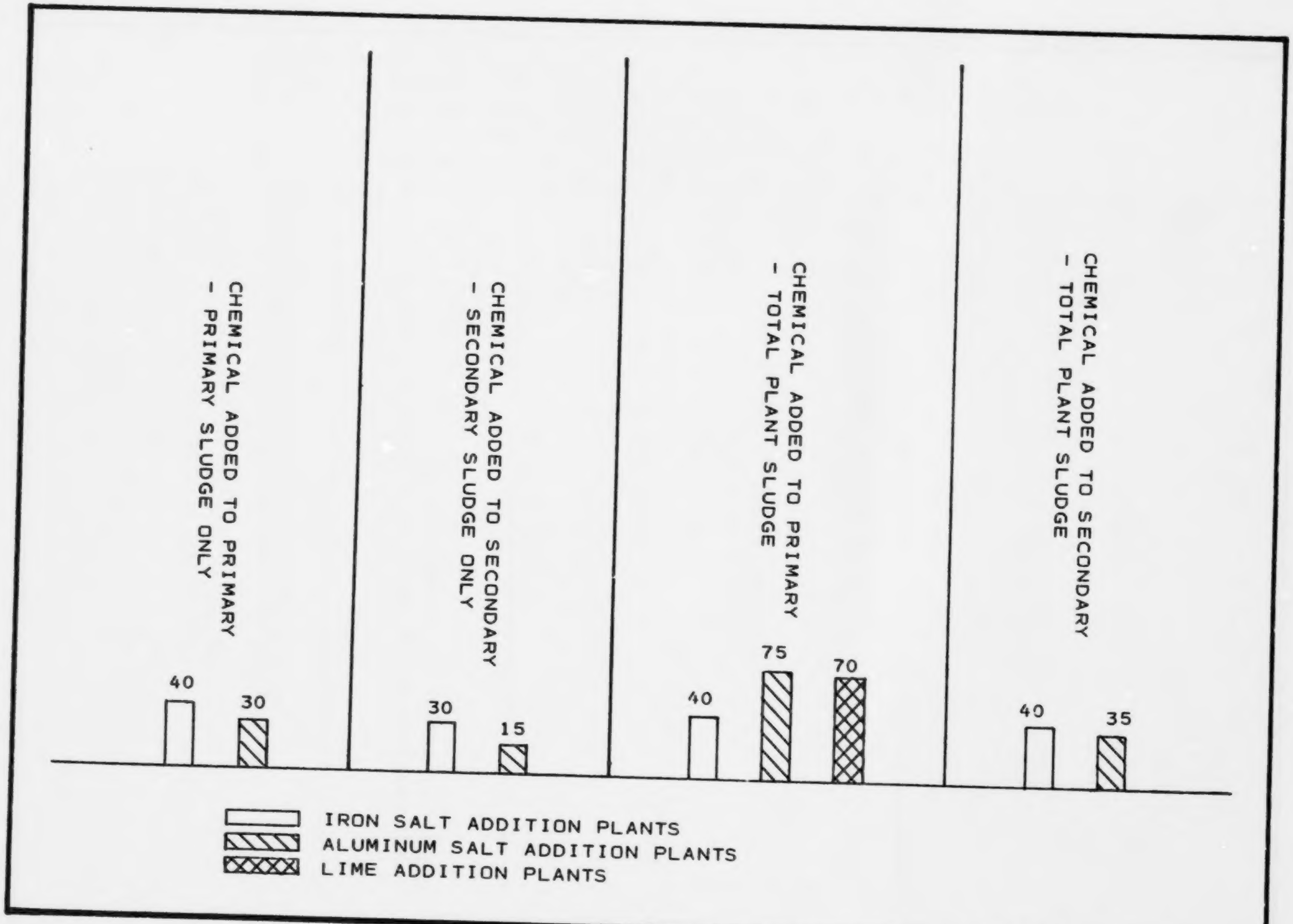


Figure 4. Kilogram (kg)sludge per kg phosphorus removed.

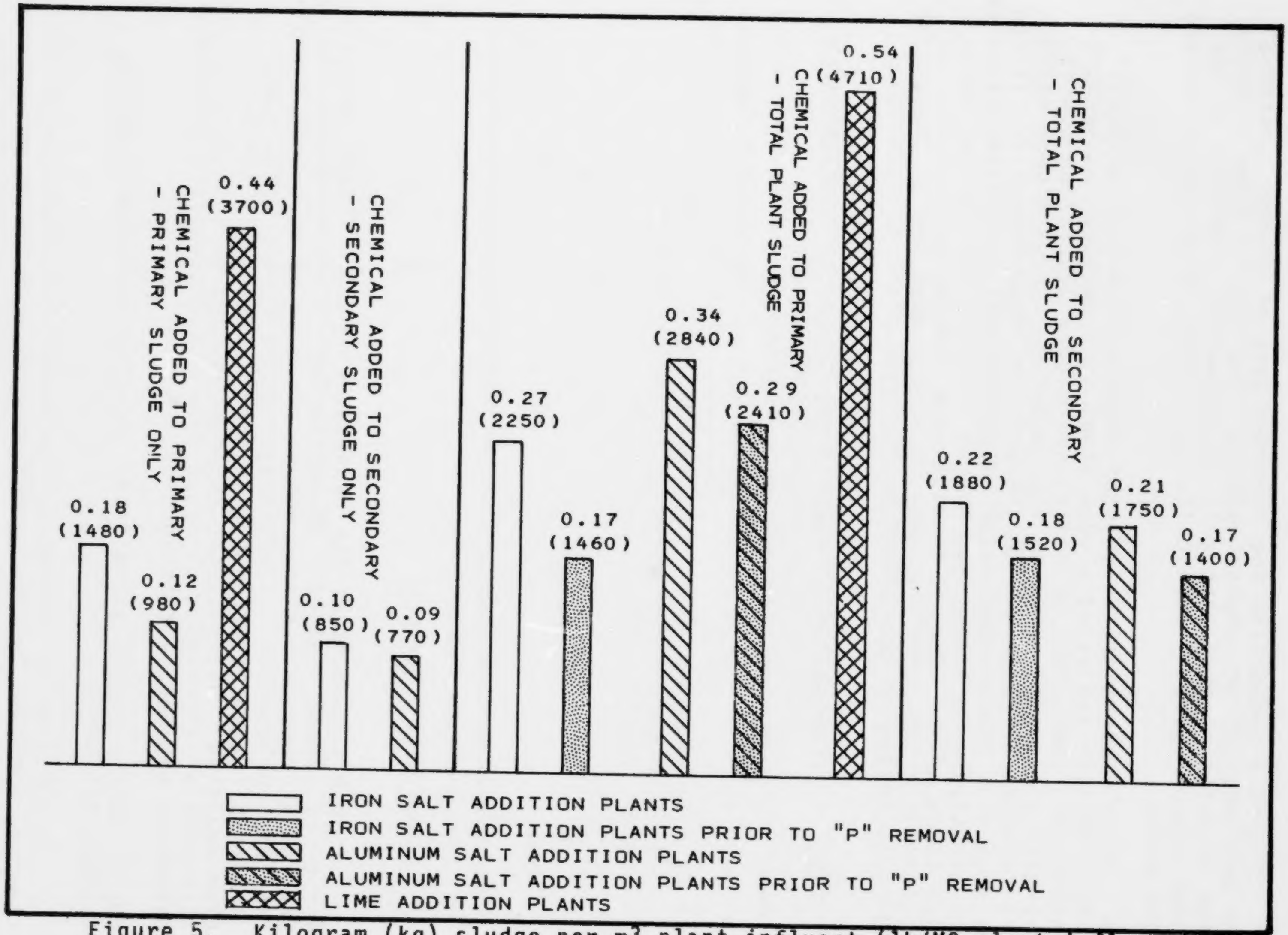


Figure 5. Kilogram (kg) sludge per m3 plant influent (lb/MG plant influent).

Figure 4 summarizes the reported mass of sludge generated per unit mass of phosphorus removed. The results were inconclusive in defining differences in sludge weight generated by the three common types of chemicals, i.e., lime, aluminum salts, and iron salts.

Table 7 shows the calculated weights of solids generated at a hypothetical activated sludge plant with addition of various chemicals. The effect of additional sludge generation due to improved plant efficiency is not included. These sludge weights are compared in Table 8 with the averages from Figure 4. It can be seen that the average rates of sludge mass generation based on the questionnaire survey are higher than the theoretical quantities calculated. This is a reasonable finding because the survey data reflect the total increase in sludge mass due to phosphorus removal, including the effect of improved plant efficiency.

TABLE 8. COMPARISON OF THEORETICAL SOLIDS GENERATION RATES WITH RESULTS OF QUESTIONNAIRE SURVEY

Chemical Added	kg of dry solids/kg of P removed	
	Theoretical *	Questionnaire†
Lime	60	70
Aluminum	34	35 - 75
Iron	36	40

* Derived from Table 7.

† Taken from Figure 4.

Figure 5 summarizes the reported weight of sludge generated per m^3 of sewage treated. Of particular interest on this figure are the increases in sludge weight reported after chemical addition was implemented. The average weight of the total plant sludge increased 54 percent with iron addition, 18 percent with alum addition. Overall, the reported sludge weights were on the high side of the ranges expected from the theoretical calculations presented earlier in this section.

SOLIDS CONCENTRATION AND PERCENT VOLATILE SOLIDS OF CHEMICAL SLUDGES GENERATED

In addition to changes in the amounts of sludges generated, chemical addition for phosphorus removal also has an effect on sludge physical characteristics. The major impacts on sludge characteristics, no matter which chemical is added for phosphorus removal, are on the sludge total and volatile solids concentrations. Table 9, which is based on the questionnaire survey of plants practicing phosphorus removal, shows the solids characteristics of the sludges before thickening or other processing. The sludges produced by iron primary addition can be seen to have had the greatest average TS concentrations. Somewhat less concentrated were the sludges produced by iron tertiary and alum primary addition. The average TS concentrations of the iron tertiary addition and alum primary addition sludges were similar. The least concentrated sludges were those produced by iron and alum addition to the secondary step. The average TS concentrations of the iron secondary addition and alum secondary addition sludges were similar. The table indicates that for all of the iron and alum sludges, the average sludge TS concentration tended to be greater when the chemical sludge was combined with the other sludges produced in the plant.

The sludge VS fraction (as a percentage of the TS concentration) appears to have been highest when alum was added to the secondary step. Alum sludges had higher VS fractions than iron sludges; and sludges produced by secondary iron or alum addition had higher VS fractions than those produced by primary or tertiary addition. The table shows that iron-tertiary sludge had a very low VS fraction; but that, when it was combined with the other sludges produced in the plant, the total plant sludge had a normal VS fraction.

The information in the table shows that the sludges produced with lime had low TS concentrations. This information could be misleading, however, unless it is remembered that the sludges were sampled before thickening. Lime sludges are readily thickened to around 10 percent TS. The VS fraction of the lime-tertiary sludge was low, even when combined with other plant sludges. Unfortunately, no information on the volatile contents of sludges produced by primary addition of lime was contained in the questionnaire.

TABLE 9. SOLIDS CHARACTERISTICS OF CHEMICAL SLUDGES WITH AND WITHOUT COMBINATION WITH OTHER PLANT SLUDGES

Type of chemical sludge and whether combined with other plant sludge(s)	Sludge characteristics (before thickening, digestion, etc.)			
	Total solids (%)		Volatile solids (% of TS)	
	Range	Average	Range	Average
Iron addition to primary step -				
Primary sludge	3.4 - 8.0	5.26	45-69	57
Total plant sludge	2.31-10.0	5.73	40-70	57
Iron addition to secondary step -				
Secondary sludge	0.2 - 4.0	0.93	50-70	62
Total plant sludge	0.5 - 7.75	4.13	42-72	62
Iron addition to tertiary step -				
Tertiary sludge	4.0	4.0	35	35
Total plant sludge	4.64- 5.0	4.82	62	62
Alum addition to primary step -				
Primary sludge	3.3 - 4.35	3.95	61-67	65
Total plant sludge	3.96- 5.0	4.49	46-70	59
Alum addition to secondary step -				
Secondary sludge	0.4 - 4.4	1.41	60-78	67
Total plant sludge	1.0 - 7.0	3.82	52-70	59
Lime addition to primary step -				
Primary sludge	0.7 - 1.5	1.1	N/A	N/A
Total plant sludge	0.64- 0.82	0.73	N/A	N/A
Lime addition to tertiary step -				
Tertiary sludge	2.5 - 4.0	3.3	11-30	21
Total plant sludge	1.95	1.95	39	39

Note: N/A = not available.

SECTION 4

PREVALENCE OF VARIOUS TREATMENT AND DISPOSAL METHODS FOR CHEMICAL SLUDGES

INTRODUCTION

The sludges resulting from chemical addition for phosphorus removal at municipal wastewater treatment facilities generally require treatment before final disposal. The many unit processes used for handling and disposal of these chemical sludges can be segregated into the following categories: sludge thickening; stabilization/reduction; conditioning/stabilization; dewatering; heat drying; reduction; and final disposal.

The prevalence of certain unit processes within the abovementioned categories, as determined from the treatment plant questionnaire survey will be discussed on the following pages. This narrative should be reviewed in conjunction with the information on Table 10.

SLUDGE THICKENING

Gravity thickening was the most prevalent thickening technique for all types of chemical sludges, and was practiced by 85 percent of the plants with a thickening step. Flotation thickening was used for thickening of waste activated sludges or occasionally for combined primary sludge and waste activated sludge. This was in accordance with the common application of air flotation to the separate thickening of waste-activated sludges. The sludges that were flotation thickened were produced with secondary addition of iron or aluminum, except in one case of primary addition of iron. Interestingly, centrifuge thickening was practiced at six plants using lime, but was not applied to the thickening of iron or aluminum sludges.

SLUDGE STABILIZATION/REDUCTION

Anaerobic and aerobic digestion were the most prevalent sludge stabilization/reduction processes. Among plants using iron salts, anaerobic digestion was used considerably more frequently than was aerobic digestion. In contrast, aerobic digestion was used almost as often as was anaerobic digestion among plants using aluminum salts. Anaerobic digestion of a lime sludge

TABLE 10. PREVALENCE OF TREATMENT AND DISPOSAL PROCESSES FOR CHEMICAL SLUDGES AMONG PLANTS RESPONDING TO QUESTIONNAIRE SURVEY *

Sludge Treatment and Disposal Unit Processes	Major Chemical Used and Point of Addition									Total
	Iron Salt			Aluminum Salt			Lime			
	Primary	Secondary	Tertiary	Primary	Secondary	Tertiary	Primary	Secondary	Tertiary	
Thickening										
Gravity	18	25	2	4	18	1	8	0	0	76
Flotation	1	4	0	0	5	0	0	0	0	10
Centrifuge	0	0	0	0	0	0	5	0	0	5
Stabilization/ Reduction										
Composting	1	0	0	1	0	0	0	0	0	2
Aerobic Dig.	4	11	1	1	24	0	0	0	0	41
Anaerobic Dig.	27	37	1	5	25	1	1	0	0	97
Lime Stabilization	0	0	0	1	0	0	0	0	0	1
Conditioning/ Stabilization										
Chemical Conditioning	11	12	1	2	12	0	2	0	0	40
Elutriation	1	2	0	0	0	0	0	0	0	3
Thermal Conditioning	2	4	0	1	4	0	0	0	0	11

(continued)

TABLE 10 (continued)

Sludge Treatment and Disposal Unit Processes	Major Chemical Used and Point of Addition									Total
	Iron Salt			Aluminum Salt			Lime			
	Primary	Secondary	Tertiary	Primary	Secondary	Tertiary	Primary	Secondary	Tertiary	
Dewatering										
Pressure Filter	1	3	0	1	1	0	2	0	1	8
Air Drying Beds	17	20	0	2	18	0	1	0	0	58
Centrifuge	0	0	0	1	4	0	0	0	1	6
Vacuum Filter	12	13	1	1	10	0	0	0	0	37
Horizontal Moving Screen Concentrator	0	0	0	0	2	0	0	0	1	3
Cylindrical Rotating Gravity Filter	0	0	0	0	1	0	0	0	0	1
Lagoon	4	4	1	1	3	1	0	0	2	16
Heat Drying										
Flash Dryer	0	0	0	0	1	0	0	0	0	1
Multiple Hearth Dryer	0	1	0	0	1	0	0	0	0	2

(continued)

TABLE 10 (continued)

Sludge Treatment and Disposal Unit Processes	Major Chemical Used and Point of Addition									Total
	Iron Salt			Aluminum Salt			Lime			
	Primary	Secondary	Tertiary	Primary	Secondary	Tertiary	Primary	Secondary	Tertiary	
Reduction										
Incineration	7	5	0	0	5	0	5	0	0	22
Final Disposal										
Agricultural Fields, Lawns, Gardens	18	39	1	7	30	1	5	0	1	102
Land Reclamation		1	0	1	6	0	1	0	0	9
Sanitary Landfill	16	18	1	1	13	0	4	0	0	53
Private- or Authority-owned Dump-site	8	10	0	2	8	0	2	0	1	31

---

[*] One plant may use more than one method of thickening, dewatering, etc.

was occurring at one plant. Three plants reported having composting operations involving iron or aluminum sludges. Chemical stabilization by lime was practiced infrequently.

SLUDGE CONDITIONING/STABILIZATION

Chemical conditioning was the most prevalent conditioning method and was used for conditioning all types of chemical sludges prior to dewatering. Thermal conditioning was in use at eleven plants which had iron or aluminum sludges. Elutriation was also practiced by three plants with iron sludges.

SLUDGE DEWATERING

Dewatering of chemical sludges on air drying beds was more common than other dewatering methods, including vacuum filtration, which followed it in frequency. Both pressure filters and centrifuges were also receiving application at a number of plants, pressure filters being used for iron, aluminum and lime sludges, and centrifuges for aluminum and lime sludges only. Horizontal and cylindrical screens were infrequently used. Sludge lagoons were common, but often they were not the sole dewatering method used at a plant.

SLUDGE HEAT DRYING

Heat drying by either flash or multiple hearth dryers was practiced in only three plants.

SLUDGE REDUCTION

Incineration with multiple hearth or fluidized bed incinerators was practiced at 22 plants. Most of these plants handled iron sludges, but both aluminum and lime sludges were also incinerated.

SLUDGE FINAL DISPOSAL

The two major final disposal sites for chemical sludges were agriculture fields, lawns, or gardens; and sanitary landfills or dumps. A sanitary landfill is an engineered operation involving the spreading of refuse on land in thin layers which are compacted and covered with earth each day. In contrast, a dump is a state-approved site where sludge is spread or buried on land (not to be later removed) and is not as closely controlled as a sanitary landfill; neither does it receive refuse.

SECTION 5

THICKENING OF CHEMICAL SLUDGES

GRAVITY THICKENING

Introduction

As previously shown in Table 10, a total of 76 (44 percent) of the plants responding to the questionnaire survey maintained gravity thickeners. Among the plants which reported having a thickening step, 85 percent utilized gravity thickening. Gravity thickening is simple and is the least expensive of the available thickening processes used for sludge concentration prior to digestion and/or dewatering. The process allows blending of sludges and sludge flow rate equalization, thereby improving the uniformity of feed to the subsequent processes.

Questionnaire Survey

Nine plants which were surveyed indicated that chemical addition was having a significant impact on their gravity thickening process. The impacts were attributed to both increased sludge quantities and changed sludge characteristics and ranged from improvements in thickening to adverse impacts. The experiences of five of these plants are summarized in Table 11. Four other plants were case study sites and will be discussed further on.

Sludge settleability was a problem at two of the three plants using iron; the third plant experienced improved settleability during iron addition. The two plants using lime found that the sludge thickened readily to 7 to 10 percent TS. Insufficient thickener capacity for the additional sludge was a concern at one of these plants, however.

Case Studies

The case studies in the appendices contain detailed information on the effects of chemical addition on gravity thickening. The particular case studies which deal with this subject are Sheboygan, Wisconsin; Port Huron, Michigan; Lakewood, Ohio; and South Bend, Indiana. The experiences at these plants are similar to the experiences of plants responding to the questionnaire survey. However, some unusual highlights of the case studies can be mentioned.

TABLE 11. IMPACTS OF CHEMICAL SLUDGE UPON GRAVITY THICKENING PERFORMANCE AS REPORTED IN QUESTIONNAIRE RESPONSE

Location	Size m^3/day (mgd)	Type of Raw Sludge Treated	Impacts of Chemical Sludge
Milford, Michigan	3,000 (0.8)	Iron-secondary and primary from a conventional activated sludge plant	The sludge thickens better and is forming more supernatant. There is more room for sludge in the thickeners, which are operated as holding tanks.
Wayne Co.-Wyandotte, Michigan	291,400 (77)	Iron-primary and secondary from a pure oxygen activated sludge plant	The sludge is harder to thicken. It is lighter (fluffier).
Ford Lake WWTP, Ypsilanti, Michigan	22,700 (6)	Iron-secondary and primary from a step aeration activated sludge plant	It has been very difficult to handle the extra sludge generated by phosphorus removal because the thickener bulks back to the aeration tanks before wasting is finished.
Virginia, Minnesota	8,700 (2.3)	Lime-primary and secondary from a conventional waste activated sludge plant	The plant converted from anaerobic digestion and drying beds to gravity thickening, lime stabilization and lagoons in anticipation of phosphorus removal with lime. The thickener produces an 8 to 10 percent TS sludge.
Hatfield Twp.-Colmar, Pennsylvania	8,700 (2.3)	Lime-primary, aluminum-tertiary and secondary from a complete mix activated sludge plant	The sludge thickens to 7.0 percent TS. Although sludge settleability is fine, the additional sludge is overloading the thickener. The thickener overflow contains 8,400 mg/ℓ SS and 3,500 mg/ℓ BOD. The superintendent feels that the high solids content of the overflow is not hurting the plant because the associated aluminum mixes with the raw sewage and lime and may improve primary clarification.

At the Sheboygan, Wisconsin, case study site, primary and secondary sludges were fed to a single thickening unit and gravity thickened to an average TS concentration of 8.6 percent before chemical addition for phosphorus removal began. After secondary addition of ferric chloride was incorporated into the treatment system, the sludge was gravity thickened to 7.76 percent TS on the average. The thickened sludge solids concentration was lower after phosphorus removal was started despite an increase in the average concentration of the primary sludge fed to the thickener. Problems with floating of the sludge blanket and resulting high solids overflow were experienced in thickener operation both before and after phosphorus removal, but were more frequent after phosphorus removal. The problems with high floating sludge tended to coincide with industrial waste discharges. A means of reducing thickener solids overflow was found which involved adding polymer separately to the primary and secondary sludge feed lines to achieve the same concentration of polymer in the sludge in each line. In other words, the amount of polymer added to each line was proportional to the sludge flow. The polymer was added through a plastic hose to a point immediately before the discharge end of the feed lines into the thickener center well. The usual dosage was 3 ppm. During the first 4 months after proportional polymer addition was started, the thickener overflow TS concentration averaged only 154 mg/ℓ, compared to previous concentrations as high as 2,000 mg/ℓ TS.

At Port Huron, Michigan, aluminum sulfate was added to the activated sludge basins for phosphorus removal. The plant used a Dorr-Oliver "Densludge" type SD gravity thickener. Waste activated sludge was pumped to a box ahead of the thickener where it was combined with primary sludge and polymer. On the average, the raw sludge thickened from 0.56 to 4.68 percent total solids, and the overflow contained approximately 2,480 mg/ℓ SS. However, due to variations in the activated sludge wasting rate and intermittent incinerator operation, performance of the gravity thickener was very inconsistent.

The thickened sludge concentration reached almost 6 percent on a few good days; only about 4 percent on poor days. The overflow solids concentration varied greatly from day to day between about 100 mg/ℓ and 10,000 mg/ℓ. These variations were related to the addition of alum to the aeration basins. Alum addition increased the mass of activated sludge which was generated and wasted. This waste activated sludge had poorer thickening characteristics than the primary sludge, so that when activated sludge wasting rates were high, thickening was poorer. Low thickened sludge solids concentrations led to higher polymer requirements for chemical conditioning.

Recently at Port Huron, modifications in thickener operation have been made with significant results. Two identical gravity thickeners are used rather than just one. In the past, the sludge blanket depth approached 2.1 m (7 ft) in the single 3.4

m- (10 ft-) deep thickener. Polymer addition was necessary to aid solids capture and thickening. With two thickeners, the sludge blanket depth is maintained between 1.2 and 1.5 m (4 and 5 ft), and no polymer is used.

The savings in polymer cost amounts to \$2.75/t (\$2.50/ton) of dry solids. It is also reported that a reduction in the amount of phosphorus recycled to the head of the aeration basins in the thickener overflow has occurred. This may mean a slight decrease in the cost of phosphorus removal chemicals.

The Lakewood, Ohio, plant treats aluminum waste activated and primary sludges which are combined in gravity thickeners. The thickener overflow is returned to the head of the plant. Lakewood is an older facility whose effluent quality is impaired due to hydraulic overloading and inadequate solids handling capacity. Alum addition adversely affected gravity thickener operation as the result of the generation of additional sludge solids. The mass of waste activated sludge pumped to the thickeners increased by 4.3 t/mil m^3 of (360 lb/MG) of wastewater treated. These additional solids increased the overloading of the thickeners, and problems with bulking and poor overflow quality became more severe. The thickener overflow quality deteriorated to the extent that the overflow contributed a higher loading of SS to the primary clarifiers than did the plant influent. The plant alleviated the problem by removing sludge from the thickeners at a faster rate by increasing the capacity for digestion, dewatering, and final disposal. Inadequate capacity in these areas had been causing a bottleneck in the solids-handling system.

At South Bend, Indiana, lime was added to two tertiary upflow clarifiers for phosphorus removal. The volume of lime sludge generated was 1,158 m^3/day (306,000 gal/day). The average sludge TS concentration was 4.0 percent. Gravity thickening reduced the sludge volume to 454 m^3/day (120,000 gal/day) and concentrated the sludge to 10.2 percent TS.

Literature

Accounts of gravity thickening of chemical sludges presented in the literature are summarized in the following paragraphs.

An EPA pilot plant using lime for phosphorus removal and lime recovery by recalcination found that the recycling of the lime markedly improved the thickening properties of the sludge (5). With lime recycling, the solids were reduced to 26 percent of their original volume after one hour of settling. Without lime recycling the same volume reduction required approximately 5 hours. Typically the gravity thickener produced sludges of 15 to 20 percent solids for either sludge type.

For a 15 percent solids thickener underflow concentration, the improved settling characteristics increased the thickener capacity from a loading rate of 92 $kg/m^2/day$ without lime recycling to 1,470 $mg/m^2/day$ with recycling. Since the increase in thickener capacity was much greater than the actual increase in solids loading, less thickener area was required during lime recycling.

The Holland, Michigan, wastewater treatment plant adds lime to the primary clarifiers for phosphorus removal (43). Excess activated sludge from the final clarifiers is pumped to the thickener and combined with the primary sludge. An adequate supply of dilution water is mixed with the activated sludge fed to the thickener. The dilution water, clarified secondary effluent, serves two purposes. It contains dissolved oxygen which helps keep the thickener sludge from becoming septic. It also lowers the solids concentration in the feed to a level where the particles are not hindered in settling. The thickener has produced an underflow solids concentration of 7.4 to 17 percent TS.

To improve thickener operation at Holland, the following modifications were made: The thickener was provided with vertical pickets to help release supernatant from the sludge in compression; additional vertical mixing bars were attached to aid this process; and there was a modification to the rakes to prevent furrowing the sludge blanket.

The centrate from the sludge centrifuges was originally returned to the sludge blanket in the thickener. The return of centrate to this blanket disrupted both the thickening and clarification process. The centrate lines were re-routed to the ash thickener. The overflow from the ash thickener was returned to the plant influent.

An EPA study by the Wastewater Research Division of the Municipal Environmental Research Laboratory (76) examined the thickening characteristics of an aerobically digested aluminum sludge with and without polymer conditioning. Aerobically digested waste activated sludge grom a 2 to 3 mgd plant was thickened in a 2-ℓ graduated cylinder. The results were poor: At initial sludge TS concentrations of greater than 1 percent, virtually no thickening occurred. With an initial sludge TS concentration of 0.86 percent, thickening to 1.62 percent was possible. Polymer conditioning improved thickening somewhat, but the final sludge TS concentration was still below 3 percent. Significant thickening occurred only at an initial TS concentration of less than 1 percent both with and without polymer conditioning.

At Portage Lake, Michigan, gravity thickening of an aerobically digested alum sludge was measured in laboratory cylinders (2).

(See page 64 for a description of the Portage Lake plant.) Waste activated sludge produced alternatively with and without alum addition to the plant's aeration basins was digested in full-scale aerobic digesters. The laboratory tests showed that the sludge produced with alum addition thickened more rapidly and to a higher solids concentration than the sludge produced without alum addition.

Conclusions

The impacts of chemical addition for phosphorus removal on sludge gravity thickening are attributable to both increased sludge quantities and changed sludge characteristics.

Lime addition for phosphorus generally improved the thickening characteristics of the sludge. Gravity thickened lime sludges typically have an average TS concentration of 10 percent, but the approximate range is 7 to 20 percent. In contrast, sludge thickening characteristics may be either favorably or adversely affected by the addition of iron or aluminum salts. At each plant, the impact will depend on the influent wastewater characteristics, the type of wastewater treatment, the relative proportions of primary and secondary sludges, and the sludge pumping procedures.

Thickener overloading is a frequent result of the increased amounts of sludge generated with any of the phosphorus removal chemicals. Even slight overloading causes serious problems because of the sensitivity of gravity thickeners to the critical control variables of loading rate and sludge blanket depth. In most plants where problems have occurred, however, thickener capacity was already insufficient before phosphorus removal began.

The information which has been presented suggests several ways of modifying plant operation to overcome some of the adverse impacts of chemical sludges on thickening. For instance, adding lime or a polymer to the sludge as it enters the thickener can be helpful. At Sheboygan, Wisconsin, it was found that the success of polymer addition depended upon finding the right point of addition and dosage rate through a considerable amount of experimentation.

At several of the case study sites, return sidestreams were suspected of disrupting thickener operation. Recycle streams such as digester supernatant and centrifuge centrate can carry heavy loads of difficult-to-settle solids which have a negative impact on sludge settleability. Treating these sidestreams or rerouting them to a different point in the plant may indirectly cause an improvement in thickener operation.

FLOTATION THICKENING

Introduction

Flotation thickening is becoming an increasingly popular method of sludge thickening prior to further handling. Among the variables that affect flotation thickening performance are:

- Sludge feed solids concentration
- Sludge detention period
- Type and quality of sludge
- Solids and hydraulic loading rates
- Use of chemical aids.

As a general rule, the higher the solids loading rate, the lower the solids concentration of the thickened sludge. When thickening a mixture of primary and waste activated sludge, a higher solids loading rate is allowed and a thicker sludge is produced than when thickening waste activated sludge alone. Higher loading rates and thicker sludges also result from chemical aids (usually cationic polyelectrolytes), which increase the capture of solids, improving the quality of the subnatant.

Chemical addition for phosphorus removal can affect process performance by changing the relative amounts of primary and waste activated sludge generated in a plant, by altering feed sludge solids concentrations, or by modifying feed sludge quality. Chemical addition can also affect process economics by changing the relative amounts of primary and waste activated sludges generated. On an economic basis, flotation thickening is considered to be most applicable to waste activated sludges, and is generally used when a plant has no primary sludge or when separate gravity thickening of primary sludge is provided.

Questionnaire Survey

Used by 6 percent of the plants responding to the questionnaire survey, flotation was shown to be the second most common method of chemical sludge thickening. Unlike the more common gravity thickening, which is applied to both trickling filter and waste activated sludges, flotation is applied only to waste activated sludge. Nine plants which responded to the questionnaire survey commented upon the performance of their flotation units and the impacts of the chemical sludges. Table 12 summarizes these comments, indicating the type of sludge treated at each plant and the type of sludge treatment system.

All but one of the plants treated iron or aluminum waste activated sludges. These sludges were produced by secondary addition of the chemical and were not mixed with primary sludge before thickening. The other plant was an exception because it practiced primary chemical addition and it thickened a mixture

TABLE 12. IMPACTS OF CHEMICAL SLUDGES UPON FLOTATION THICKENER PERFORMANCE AS REPORTED IN QUESTIONNAIRE RESPONSE

Plant Location	Type of Sludge Treated	Impacts of Chemical Sludge	Performance with Chemical Sludge	Further Treatment
Warren, Michigan	Alum-secondary (aerobically digested)	Although the feed sludge solids concentration remained the same, it was necessary to reduce the hydraulic loading rate by 10 or 20 percent. No change in polymer dosage was necessary.	Sludge at 1 percent TS is thickened to 5 percent.	Blending with primary sludge; Polymer conditioning; Vacuum filtration; Incineration
Zilwaukee, MIchigan	Iron-secondary (aerobically digested) from a plant with no primary treatment	The plant has always treated iron sludge and, therefore, comparisons cannot be made.	The sludge from the aerobic digester averages only 1.5 percent TS. It was thickened by flotation to only 2.5 percent TS. Flotation was discontinued because it could not deliver sludge fast enough for the vacuum filters. Polymer as a thickening aid was tried and it lowered the SS concentration of the subnatant from 50 to 12 mg/ℓ, but had no other effect.	Polymer conditioning; Vacuum filtration; Sanitary landfill
McHenry, Illinois	Iron-primary and secondary (aerobically digested)	The SVI of the chemical sludge is only 60 to 70 compared to a high SVI before iron addition The proportion of waste activated sludge in the thichener feed has increased. Before iron addition, cationic polymer was used as a thickening aid and the sludge was easily concentrated to 7 to 8 percent TS. With the iron sludge it has been necessary to use both anionic and cationic polymers and the thickened sludge only reaches 5 to 6 percent TS.		Polymer conditioning; Vacuum filtration; Croplands

(continued)

TABLE 12 (continued)

Plant Location	Type of Sludge Treated	Impacts of Chemical Sludge	Performance with Chemical Sludge	Further Treatment
Gurnee, Illionis	Iron-secondary	No impacts were noticed.	The feed sludge at 8500 ppm SS is thickened to 4.8 to 5.1 percent TS. The subnatant contains 70 to 80 mg/ℓ SS. Polymer is used to aid solids capture.	Combined with primary sludge; Lime and $FeCl_3$ conditioning; Vacuum filtration; Sanitary landfill
Oak Creek, Wisconsin	Iron-secondary	No impacts were noticed.	Sludge at 1.0 to 1.5 percent TS is thickened to 3.5 to 4.0 percent. The subnatant SS concentration ranges under 500 ppm. Polymer is used when a high loading rate is needed to keep up with a high activated sludge wasting rate.	Combined with primary sludge; Anaerobic digestion; Drying lagoons; Croplands or sanitary landfill
Willoughby-Eastlake, Ohio	Alum-secondary	No impacts were noticed.	The sludge can be concentrated to 4 percent, or as high as 7 percent. At the higher concentration, the sludge loading rate is slower and the subnatant contains more solids. The subnatant SS concentration varies from 52 to 3000 mg/ℓ. Polymer to aid thickening is not used.	Combined with primary sludge; Lime and $FeCl_3$ conditioning; Vacuum filtration; Incineration or dumpsite
Middleburg Hts., Ohio	Alum-secondary (aerobically digested) from plant with no primary treatment	The average thickened sludge TS concentration was 2.5 percent before alum addition and since has been 3.3 percent TS.	The subnatant SS concentration is averaging 300 mg/ℓ SS.	Lime and $FeCl_3$ conditioning; Vacuum filtration; Incineration

(continued)

TABLE 12 (continued)

Plant Location	Type of Sludge Treated	Impacts of Chemical Sludge	Performance with Chemical Sludge	Further Treatment
Kenosha, Wisconsin	Iron-secondary	No impacts were noticed	Sludge feed at 1 percent TS is thickened to 4 percent TS. No polymer is used to aid thickening because it seems to provide no significant improvement.	Combined with primary sludge; Anaerobic digestion; Lime and $FeCl_3$ conditioning; Pressure filtration; Croplands
Frankenmuth, Michigan	Primary sludge is anaerobically digested then sent to activated sludge aeration basins. The aluminum-secondary sludge is flotation thickened	The plant has tried iron for phosphorus removal but feels that the sludge is easier to thicken and dewater when sodium aluminate is used.	Feed sludge at 0.6 percent SS can be concentrated to 3.5 percent TS, or as high as 4.0 percent. At the higher concentration the loading rate is slower, however. Thickening is easiest when the sludge density index of waste activated sludge is kept above one. A small concentration of polymer is used (.011 kg/kg of solids) to help the sludge fall off the skimmers neatly, but it does not increase the sludge solids concentration. The subnatant averages 209 mg/ℓ SS and 100 mg/ℓ BOD.	Lime and $FeCl_3$ conditioning; Vacuum filtration; Sanitary landfill

of primary and secondary sludges. Aerobic digestion of the sludge before thickening was practiced by half of the plants listed in the table, while fewer anaerobically digested after thickening. All nine plants had a dewatering step, most using vacuum filters, but one using a pressure filter and another using drying lagoons.

Four of the plants reported a significant change in flotation unit performance related to the chemical sludges, while four reported no change at all. A ninth plant could not comment on the impact of chemical addition, but described the performance of their flotation unit with a chemical sludge. Among the four plants that noted significant changes, one reported a positive impact: an increase in the thickened sludge TS concentration. The other three plants reported negative impacts: the need to reduce the sludge hydraulic loading rate, the need to use both anionic and cationic polymers as thickening aids, and a decrease in the thickened sludge TS concentration.

Some of the plants reported that the addition of a polymer to aid thickening was not necessary; others found that polymer addition increased the allowable sludge loading rate or improved the subnatant quality. One plant noted that thickener performance was aided by running the bottom flights of the unit continually instead of just when the bottom pumps were on. Two plants related good performance to either a high sludge volume or sludge density index.

Literature

A theoretical discussion of sludge generation rates by Knight, et al. (40), indicated that the waste activated sludge portion of the total sludge produced in a plant is changed when a phosphorus removal chemical is added ahead of the primary clarifiers. Theoretically, in a conventional activated sludge plant, 44.5 percent of the total sludge mass is waste activated; in the same plant with primary chemical addition, only 8 to 14 percent of the total sludge mass is waste activated. This is because of the increased SS and BOD removals that occur during primary treatment when chemicals are added. According to the authors, because only 18 to 14 percent of the sludge is waste activated, separate flotation thickening of waste activated sludge is less likely to be economical in a plant with primary chemical addition than in a conventional plant or one with secondary or tertiary addition. The advantages of flotation thickening would be diminished with such a small percentage of waste activated sludge.

At Portage Lake, Michigan, the flotation thickening capacity of an aerobically digested alum sludge was measured in batch laboratory-scale tests (2). (See page 64 for a discussion of the Portage Lake plant.) Waste activated sludge produced alternatively with and without alum addition to the plant's aeration basins was digested in full-scale aerobic digesters. The laboratory-scale tests indicated no difference between the thickening properties of the sludges produced.

Conclusions

In some cases, chemical sludges may flotation thicken just as well as, or better than, regular sludges. However, in other cases, chemical sludges require a lower sludge loading rate or different polymers, and lower concentrations of thickened sludge solids are achieved. Little information is available for drawing further conclusions about the flotation thickening properties of various chemical sludges. It is necessary that further information be collected, however, in order to make judgements about the cost effectiveness of flotation thickening of chemical sludges. In particular, information on allowable solids loading rates with chemical sludges is needed. When the cost effectiveness of flotation thickening is compared to that of gravity thickening, a higher sludge loading rate is frequently one of the most important factors in favor of flotation.

It is also necessary to determine allowable loading rates for chemical sludges so that design engineers can size equipment properly. It seems that presently, equipment is often undersized. At some of the plants listed in Table 12, a low sludge throughput rate was a problem. The plants were having trouble achieving a high rate that would avoid backlogs of sludge in the clarifiers, and that would provide thickened sludge fast enough for subsequent dewatering operations. The plants also preferred a high rate so that they could run their sludge handling equipment for only one shift per day.

SECTION 6

STABILIZATION OF CHEMICAL SLUDGES

ANAEROBIC DIGESTION

Introduction

As previously shown in Table 10, 97 (56 percent) of the plants responding to the questionnaire survey reported that they have anaerobic digestion. As with other unit processes, it is necessary to evaluate the anaerobic digestion process as a component of the entire sludge treatment/disposal system. For example: In order to properly compare the costs of anaerobic digestion of regular vs. chemical sludges, differences in dewatering, disposal, and supernatant treatment costs must be analyzed for the total sludge management system.

Questionnaire Survey

Twenty-one plants which were surveyed indicated that chemical addition was having a significant impact on their anaerobic digestion process. Table 13 summarizes the experiences of those plants. All but one of the plants were treating iron and alum sludges produced by primary or secondary addition of the chemical. A single plant treated a lime sludge. At this plant there were reportedly no adverse impacts of this type of sludge on anaerobic digestion, but a slight increase in digester pH was noted.

As the table shows, the most common impact of chemical addition at the plants was a sudden increase in raw sludge volume and mass. A number of plants also noted changes in the settling characteristics of the sludge. The common results of these impacts were:

- Increased energy requirements for sludge mixing, pumping and heating
- Difficulty in achieving adequate digester mixing and heating
- Increased labor requirements for sludge pumping
- Increased digester gas production

TABLE 13. IMPACTS OF CHEMICAL SLUDGES UPON ANAEROBIC DIGESTER PERFORMANCE AS REPORTED IN QUESTIONNAIRE RESPONSE

Location	Size- m^3/day (mgd)	Type of Sludge Treated	Impacts of Chemical Sludge	Performance with Chemical Sludge
Richardson, TX	6,720 (1.8)	Alum-secondary and primary from a trickling filter plant	The raw sludge mass increased by about .033 kg/m^3 (275 lb/MG). The volume of digested sludge increased by about 50 percent. In the past, when alum was added before the primary clarifiers, the solids in the digester units stratified and upset the digestion process. When the point of addition was moved to the secondary stage, no digestion problems were observed.	The digested sludge averages 5 percent TS disposal by vacuum filtration and incineration.
Ashland, WI	4,542 (1.2)	Alum-secondary and primary from a step-aeration activated sludge plant	The plant began secondary treatment and vacuum filtration at the same time that phosphorus removal was begun. At the time of only primary treatment, digester supernatant characteristics were good. With secondary treatment and phosphorus removal, however, solids-liquid separation in the secondary digesters no longer occurred. Adding polymer to the digester did not solve the problem.	Digested sludge vacuum filtered and disposed of in a landfill.

(continued)

TABLE 13 (continued)

Location	Size-m^3/day (mgd)	Type of Sludge Treated	Impacts of Chemical Sludge	Performance with Chemical Sludge
Parry Sound, Ontario	3,220 (0.85)	Iron-primary from a plant with no secondary treatment; gravity thickening before digestion	The sludge is more difficult to digest. It is necessary to add 113 kg/mo (250 lb/mo) of lime to increase the alkalinity in the digester.	Digested sludge averages 8 percent TS. Sludge dried on drying beds and disposed of in a landfill.
Cedarburg, WI	4,542 (1.2)	Alum-secondary and primary from a conventional activated sludge plant; gravity thickening before digestion	The raw sludge volume increased. The pH in the digester was raised to 8.2. The volume of digested sludge increased, raising disposal costs.	Digested sludge is applied to drying beds or lagoons, or hauled directly to croplands.
Fergus, Ontario	2,270 (0.6)	Iron-secondary and primary from a conventional activated sludge plant	Foaming in the digesters occurred. The volume of digested sludge was increased, raising disposal costs.	Digested sludge at 5 percent TS is applied to croplands.
Three Rivers, MI	4,730 (1.25)	Alum-primary or lime primary and secondary from a conventional activated sludge plant	With alum, increased raw sludge volume and mass. Poorer solids-liquid separation in secondary digester; therefore, less supernatant can be removed. With lime, the huge volume of sludge generated plugged all the lines and filled the digesters, so it was discontinued.	Supernatant contains 2.5 percent TS and 1,920 mg/ℓ BOD. Digested sludge TS concentration is 5.3 percent.

(continued)

TABLE 13 (continued)

Location	Size-m^3/day (mgd)	Type of Sludge Treated	Impacts of Chemical Sludge	Performance with Chemical Sludge
Watt's Creek STP, Shirley's Bay, Ontario	18,620 (4.92)	Alum-secondary and primary from a conventional activated sludge plant	The raw sludge volume increased 65 percent or more as the percent solids decreased from 3.8 to 3.0 percent TS. The sludge mass increased by about 1,770 kg/day (3,900 lb/day). The retention time in the digesters was reduced, resulting in digester failure. Supernatant TS concentration increased. Raw sludge volatile content decreased from about 70 to 64 percent. Digested sludge TS decreased from 5 or 6 percent to 3 percent with an increase in volatile matter from 50 to 55 percent. The digester temperature is hard to maintain due to insufficient heat exchanger capacity with the increase in sludge volume. Additional labor costs for raw and digested sludge pumping are estimated at $32,500/yr. Several experimental changes in operating procedures are planned: use of iron for phosphorus removal with primary addition to	Supernatant contains 2.5 percent TS.

(continued)

TABLE 13 (continued)

Location	Size- m^3/day (mgd)	Type of Sludge Treated	Impacts of Chemical Sludge	Performance with Chemical Sludge
Watt's Creek STP, Shirley's Bay, Ontario (cont'd)			hopefully produce a more compact sludge; settling of supernatant in a holding tank; liming of primary sludge to improve settleability; operation as a contact stabilization plant.	
Gladstone, MI	2,839 (0.75)	Alum-secondary and primary from a bio-surf treatment plant	Increased volume and mass of sludge requires additional labor and electricity for pumping. Additional electricity is also required for mixing and natural gas for heating. Labor costs increased by \$11,400/yr (2080 labor hours). Cost of electricity for mixing sludge in digester increased by \$14.60/day (400 kwh/day). The sludge does not settle as well in the secondary digester. A heavier sludge was obtained by returning the secondary sludge to the primary clarifiers.	Both the raw and digested sludge average 4.0 percent solids. The raw sludge is 65 percent volatile.
Silver Bay, MN	2,190 (0.58)	Alum-secondary and primary (gravity thickened from a trickling filter plant)	Solids-liquid separation in the single-stage digester did not occur after alum addition.	

(continued)

TABLE 13 (continued)

Location	Size-m^3/day (mgd)	Type of Sludge Treated	Impacts of Chemical Sludge	Performance with Chemical Sludge
Ford Lake WWTP, Ypsilanti, MI	22,700 (6.0)	Iron-secondary and primary from a step-aeration activated sludge plant; gravity thickening before digestion	Increased sludge volume is crowding the digester. Digestion seems to be poorer because of decreased volatile fraction.	Raw sludge before thickening is 2.5 percent TS and 70 percent volatile. Digested sludge is 10 percent TS.
Duffin Creek WPCP, Whitby, Ontario	11,350 (3.0)	Lime-primary and iron-secondary from a step-aeration activated sludge plant	No adverse impacts on anaerobic digestion. Digester pH has increased a small amount. No additional costs were incurred.	Digested sludge TS concentration is 5.7 percent.
Welland, Ontario	34,800 (9.2)	Alum-secondary and primary from a step-aeration activated sludge plant	Both raw sludge volume and mass increased. The TS concentration of the raw sludge decreased from 5.89 to 4.89 percent. More energy was required for digester heating and sludge pumping. Sludge settling in the secondary digester appeared slower. Pumping costs increased by $6,000/yr for labor and $12,000/yr for electricity in 1976. The cost of additional natural gas for digester heating was $2,000 in 1976. The volume and mass of digested sludge increased, raising disposal costs.	The combined raw sludge averages 4.89 percent TS and the digested sludge averages 10 percent TS. Sludge disposal is by trucking liquid to croplands.

(continued)

TABLE 13 (continued)

Location	Size-m^3/day (mgd)	Type of Sludge Treated	Impacts of Chemical Sludge	Performance with Chemical Sludge
Niagara Falls, Ontario	31,600 (8.3)	Iron-primary from a plant with no secondary treatment	Raw sludge volume increased from 53 to 76 m^3/day (14,000 to 20,000 gpd) as TS concentration dropped from 9 to 7 percent. Sludge mass increased by about 454 kg (1,000 lb)/day. Digester gas production increased. The digester gas is utilized for digester and building heating, so natural gas consumption was reduced. Additional energy was required to pump sludge to and from the digesters and through the heat exchanges.	Digested sludge TS concentration averages 12 percent. Digested sludge pumped to lagoons and later trucked to croplands.
Weyauwega, WI	961 (0.25)	Iron-secondary from a conventional activated sludge plant with no primary treatment. Gravity thickening before digestion	The plant operators estimate by visual observation that the volume of raw sludge has doubled. Digestion is now more difficult. There is little or no gas production. Dewatering of the sludge by vacuum filter is more difficult. More ferric chloride and lime are used for conditioning.	The digested sludge is 5 to 6 percent TS. It is chemically conditioned, vacuum filtered to 18 to 20 percent TS and trucked to croplands.

(continued)

TABLE 13 (continued)

Location	Size-m^3/day (mgd)	Type of Sludge Treated	Impacts of Chemical Sludge	Performance with Chemical Sludge
Ottawa, Ontario	276,000 (73)	Alum-primary from a plant with no secondary treatment	Raw sludge mass increased from 4,100 to 17,700 kg/day (9,000 to 39,000 lb/day) as TS decreased from 4.8 to 4.35 percent. Raw sludge volume increased by about 90 percent. Liquid-solids separation in digester is more difficult. Increased digester gas production experienced.	Digester supernatant is 1.92 percent TS. Raw sludge feed is 67.2 percent volatile. Digested sludge is 3.5 percent TS.
Belleville, Ontario	29,900 (7.9)	Iron-secondary or alum-secondary and primary from a conventional activated sludge plant	When iron was tried for 3 months, the digester became upset; the pH dropped to 4; volatile acids production increased; gas production fell. Presently, with alum there are no problems. There was a slight increase in sludge production.	Raw alum sludge is 3.9 percent TS, 70 percent volatile. Digested alum sludge is 4.3 percent TS, 59 percent volatile. Supernatant contains 1 to 2 percent TS.
Port Dalhousie PCP, St. Catherines, Ontario	28,600 (7.55)	Iron-secondary and primary from a conventional activated sludge plant	A 30 percent increase in sludge volume resulted in an increased cost for natural gas to heat digesters.	Raw sludge is 4.1 percent TS, 62.6 percent volatile. Digested sludge is 4.3 percent TS.

(continued)

TABLE 13 (continued)

Location	Size-m^3/day (mgd)	Type of Sludge Treated	Impacts of Chemical Sludge	Performance with Chemical Sludge
Port Weller PCP, St. Catherines, Ontario	33,700 (8.91)	Alum-secondary and primary from a conventional activated sludge plant	A 26 percent increase in sludge volume resulted in an increased cost for natural gas to heat digesters.	Raw sludge is 59.6 percent volatile.
Sarnia, Ontario	30,300 (8.0)	Iron-primary from a plant with no secondary treatment	Raw sludge volume increased from 83 to 113 m^3/day (22,000 to 30,000 gpd). Sludge TS concentration decreased from 5.5 percent to 4.0 percent TS. Sludge mass increased by 36 kg/day (80 lb/day). Digester gas production increased from 1,820 to 2,240 m^3/day (65,000 to 80,000 ft^3/day). No adverse effects on anaerobic digestion were observed.	
South Lyon, MI	2,157 (.57)	Iron-secondary and primary from a conventional activated sludge plant	It is necessary to add polymer to the digester to aid settling.	Raw sludge is 7 percent TS, 62 percent volatile. Digested sludge is 6 percent TS.
Lakeview WPCP, Mississauga, Ontario	159,000 (42)	Iron-secondary and primary from a step aeration activated sludge plant	Digester pH and alkalinity were lowered. It is sometimes necessary to add lime to the digester.	The supernatant TS concentration is 1.8 percent and the BOD concentration is 6,000 to 10,000 mg/ℓ.

- Poor solids-liquid separation and high supernatant solids concentrations.

Less common results were:

- Poor digestion, accompanied by moderate decreases in digester pH, decreases in volatile matter destruction, or decreases in gas production
- Digester upsets, with symptoms such as the absence of gas production, low pH, low volatile destruction, and increased production of volatile acids
- Primary digester stratification
- Foaming in digesters.

Adverse impacts on anaerobic digestion were reported both by plants using iron and plants using alum and with both primary and secondary addition of the chemical. There was one case in which a plant experienced problems when using iron, but observed no significant adverse impacts after switching to alum. Another plant found that problems disappeared when the point of addition of alum was switched from the primary stage to the secondary stage. Some plants achieved good thickening of sludge in digesters while others achieved none at all.

Several plants reported average TS concentrations of their digested sludge. These ranged from 3.5 to 12 percent, with the average value being 6.7 percent.

The reduced volatile solids fraction of the raw sludge was felt to be significantly affecting digestion at two plants. Other plants did not report any problems resulting from the reduced VS fraction. The values for average VS fraction which were reported ranged from 59.6 to 70 percent of TS with the average value being 65 percent of TS. In many cases the VS loading was increased when the plant treated chemical sludge, and thus resulted in increased gas production per cubic meter of digester space.

Case Studies

The case studies in the appendices contain detailed information on the effects of chemical addition on anaerobic digestion. The particular case studies which deal with this subject are Lakewood, Ohio; Coldwater, Michigan; South Bend, Indiana; and Pontiac, Michigan. The experiences at those plants are similar to the experiences of plants responding to the questionnaire survey. However, some unusual highlights of the case studies can be mentioned.

At Lakewood, Ohio, anaerobic digester supernatant quantity was increased greatly when the digesters were overloaded with alum sludge. The excess supernatant was generated, not because a greater sludge volume was being fed to the digesters, but because sludge was removed from the digesters at a slower rate due to reduced vacuum filter and flash dryer capacity. The supernatant became very high in total solids. Because of the recirculation of this sidestream to the head of the plant, there were adverse impacts on the performance of both primary and secondary treatment and gravity thickener operation. By finding ways to pump more sludge out of the digester, the volume and solids concentration of the supernatant was decreased. Longer operation of the vacuum filter and flash dryer was one method used. Hauling liquid sludge was found to be a more successful method. Eventually, however, it was concluded that it would be necessary to hire a contractor to clean out the secondary digesters, removing the grit and other heavy materials which were reducing the effective volume of the digesters.

At Pontiac, Michigan, an increased volume of raw sludge was pumped to the digesters at a lower average solids concentration during phosphorus removal with ferric chloride and polymer. The net effect was an increase in the mass of primary sludge pumped to the digesters per pound of SS in the plant influent. There was an increased volume of supernatant, but at a considerably lower average solids concentration, indicative of better sludge settleability within the digesters. The chemical sludge had no negative effects on digestion. The plant manager felt that primary addition of the chemical had several benefits for plant operation which would not be realized with secondary addition: The primary sludge settled well and was easily pumped. The removal of additional solids during primary clarification because of increased clarifier efficiency with ferric and polymer use actually reduced the load on the aeration system. At times the additional primary sludge seemed to be balanced by reduced volume of waste activated sludge generated because of reduced sludge growth in the aeration basins. The increased efficiency of the primary clarifiers was very helpful in containing (and recycling) solids from the digester supernatant. If these solids were not contained they would result in the degradation of secondary treatment. The supernatant volume was always large because of inadequate sludge disposal resulting from down time on the vacuum filters. The two vacuum filters were inadequate from the start because of frequent shut-downs due to lime-scale problems. The filter media is acid cleaned once per week, but the internal piping must still be taken apart and cleaned on an annual basis. Polymer conditioning of the sludge was tried but hasn't been as effective as conditioning with lime.

At Coldwater, Michigan, the TS concentration of the sludge entering the digesters has increased since the start-up of iron and polymer addition for phosphorus removal. Most of the additional solids produced and fed to the digesters were non-volatile. The volatile fraction of the sludge decreased. But the mass of VS fed to the primary digester has increased from 590 kg VS/day (1,300 lb VS/day) before phosphorus removal to 720 kg VS/day (1,590 lb VS/day) during phosphorus removal. The amount of digester gas produced increased from roughly 5,040 m^3/mo (180,000 ft^3/mo) to 11,200 m^3/mo (400,000 ft^3/mo).

At South Bend, Indiana, combined primary and waste activated sludge was gravity thickened and then digested in two-stage anaerobic digesters. Before being fed to the primary digesters, the sludge was preheated. The plant accomplished phosphorus removal by mixing ferric chloride and polymer with the wastewater in tertiary upflow clarifiers. The resulting iron sludge was pumped from the clarifiers to a chemical sludge gravity thickener. Because of plant design, the thickened iron sludge could not be preheated before it was fed to the digesters. It therefore was capable of suppressing the temperature in the digester into which it was fed. This resulted in a loss of digester gas production. Since most of the gas production took place in the primary digesters, the chemical sludge was fed into one of the secondary digesters. The result was a 4 to 5°C loss in temperature in this digester, but a decrease in total digester gas production was avoided.

Literature

Accounts of anaerobic digestion of iron and aluminum sludges available in the literature can be summarized as follows:

- The presence of iron and aluminum precipitates in digesting sludge does not inhibit the action of the anaerobic bacteria (22, 32, 44, 51, 63).

- The volume of gas produced per pound of VS introduced to the digester is similar for conventional sludges and iron or aluminum sludges. Gas composition is also similar (42, 51).

- During anaerobic digestion of ferric iron and aluminum sludges, solubilization and release of precipitated phosphorus into the supernatant does not occur (22, 32, 44, 51, 63).

- A difference in the behavior of primary and waste activated sludges containing phosphates removed by the addition of ferrous iron occurs when they are digested. A significant release of phosphorus can take place upon anaerobic digestion of the waste activated sludges. In contrast, there is an uptake of phosphorus when the primary sludges are digested. This difference in behavior is not hard to explain. The ferrous iron added during primary treatment remains in the ferrous form before and after the digestion of primary sludge. However, the ferrous iron added to the aeration tank is mostly oxidized to the ferric form, which is then reduced back to ferrous upon anaerobic digestion, causing a release of phosphorus.

- When digesting ferric iron or aluminum sludges, high concentrations of iron or aluminum ions do not occur in the digester sludge or supernatant (44, 51).

Successful anaerobic digestion of a lime sludge was demonstrated at the 9,092 m^3/day (24 mgd) Newmarket, Ontario, activated sludge plant (6, 72). The plant achieved 80 percent phosphorus removal by adding 200 mg/ℓ lime (as $Ca(OH)_2$) to the raw wastewater The addition of lime caused the mass of primary sludge TS to increase by 0.32 kg/m^3 (2,670 lb/MG). The primary sludge containing relatively small quantities of waste activated sludge, was treated by two-stage anaerobic digestion with ultimate disposal on croplands. During lime addition, failure of the digestion process was caused by the high pH (10.0 to 10.5) of the raw sludge. Intermittent and frequent overdosing of lime was also responsible.

It was found that by letting the sludge sit in the primary clarifiers longer the pH was lowered to 9.4 and the digesters could operate satisfactorily. Modifications were made so that a sludge blanket 1.5 to 2 ft (0.46 to 0.6 m) was maintained in the clarifiers. After this modification, the digesters continued to operate satisfactorily with the primary sludge being fed at 8 to 12 percent TS compared to 3 to 4 percent TS before chemical addition. The digester operated at a pH between 7.2 and 7.4, with supernatant soluble phosphorus concentrations of 6 to 8 mg/ℓ. Digested sludge was hauled at 10 to 11 percent solids, compared to 3 to 4 percent before lime addition, and volumes were somewhat less.

The experience at Barrie, Ontario, illustrated that decreasing the dosage of the phosphorus removal chemical by 25 percent could improve sludge characteristics for digestion (72). Dosing at 200 mg/ℓ alum to the raw sewage to produce a 0.5 mg/ℓ total phosphorus concentration in the plant effluent resulted in a primary sludge concentration of 2.76 percent TS. This resulted in an increased volume of sludge being transferred to the digester, a decrease in digester temperature and gas production, and an increase in volatile acids. At an alum dosage of 150 mg/ℓ an effluent phosphorus concentration of 1.2 mg/ℓ was achieved, and the TS concentration of the primary sludge rose to 4.25 percent. This returned digester operation to normal.

Conclusions

Hydraulic or solids overloading of anaerobic digesters is the main problem to result from chemical addition for phosphorus removal. This results in inadequate liquid retention time, poor mixing, poor heating, and inadequate room for solids-liquid separation. Volatile destruction is then poor; gas production is lowered; and supernatant quality deteriorates.

The easiest remedies for these situations lie in the design stage. For instance, poor mixing is currently a major defect in the design of anaerobic digesters treating both conventional and chemical sludges. With chemical sludges, and especially lime sludges, present mixing is even more likely to be inadequate. In cases where present mixing is suspected to be inadequate, mixing equipment should be operated full time, with every advantage taken to any pipeworks and pumps which can be utilized to provide additional circulation.

The situation is similar with regard to sludge heating in digesters. Chemical sludges can aggravate existing heating problems. When there is a separate chemical sludge, it may be possible to avoid loss of digester temperature in the primary digester by introducing the chemical sludge into the secondary digester. This would prevent a suppression of digester gas production in the primary digester where most of the gas is usually produced.

Another major problem resulting from chemical addition is poor solids-liquid separation in the digester. Both iron and aluminum sludges can contain flocs which do not settle readily. Methods of improving the settling of these sludges include adding polymer to the digester and sludge liming before digestion. In some situations it may be possible to produce a sludge with better settling characteristics by changing the type of phosphorus removal chemical used, the chemical dosage, or the point of addition. Also, in some cases, using a polymer in conjunction with the iron or aluminum for phosphorus removal could improve solids-liquid separation in the digester.

When poor supernatant is a problem as the result of digester crowding, several steps can be taken to reduce the crowding. Raw sludge prethickening, when possible, can reduce sludge volume, saving digester space. Sometimes thickening can be achieved in existing holding tanks or by allowing sludge to remain on clarifier bottoms longer. Also, in some cases, recirculating waste biological sludge to the primary clarifiers will produce a thicker combined sludge.

Faster removal of sludge from digesters is another way of reducing digester crowding. Many plants may not be removing sludge at a fast enough rate because of lack of disposal ability. A popular way to meet this need is with liquid sludge hauling.

When faced with a poor supernatant, a plant may choose to avoid drawing off supernatant except at times when it is clear. However, there is then less reduction occurring in the volume of sludge for disposal. In other situations, a plant which has started phosphorus removal may be forced to remove more supernatant because of inadequate means of disposing of the additional sludge which is fed to the digesters. Settling of supernatant solids in a holding tank may then be possible to avoid the deterioration of wastewater treatment performance.

Cleaning of digesters to remove grit is expensive, but it is ultimately found to be necessary in many cases where greater space and longer retention time in digesters are needed. It is not uncommon to find that effective digester volume has been reduced drastically by accumulated grit.

AEROBIC DIGESTION

Introduction

As previously shown in Table 10, 41 (24 percent) of the plants responding to the questionnaire survey reported that they had aerobic digestion. Aerobic digestion is mostly used by smaller treatment plants and/or for treatment of waste secondary sludges. Principal advantages cited for aerobic digestion over anaerobic digestion are simpler operations, lower initial capital cost, and better supernatant quality. The major disadvantage is the much higher operating cost due to high energy consumption.

Two basic types of aerobic digestion systems exist. Single-stage digestion is a batch-type operation where the sludge is supplied with air and completely mixed for a period of time, followed by quiescent settling and decanting of supernatant, all in one tank. Two-stage digestion incorporates a separate tank for settling and supernatant decanting. Two-stage digestion is a continuous-feed rather than a batch operation.

As with other unit processes, it is necessary to evaluate the aerobic digestion process as a component of the entire sludge treatment/disposal system. For example: In order to properly compare the costs of the aerobic and anaerobic digestion systems, differences in dewatering, disposal, and supernatant treatment costs must be analyzed for the total sludge management system.

Questionnaire Survey

Four plants which were surveyed reported problems with aerobic digestion of chemical sludges. Table 14 summarizes the experiences of those plants.

The most common impact of chemical addition was a sudden increase in raw sludge volume and mass. This resulted in the need for more secondary digester capacity, the need for an increased air supply rate to achieve adequate mixing, and higher disposal costs. Hauling the liquid digested sludge to croplands

TABLE 14. IMPACTS OF CHEMICAL SLUDGES UPON AEROBIC DIGESTER PERFORMANCE AS REPORTED IN QUESTIONNAIRE RESPONSE

Location	Size- m^3/day (mgd)	Type of Sludge Treated	Impacts of Chemical Sludge	Performance with Chemical Sludge
Portland, MI	1,135 (0.3)	Iron-primary and iron-secondary from an activated sludge plant	Total raw sludge volume increased from 8 to 19 m^3/day (2,000 to 5,000 gal/day). Sludge TS decreased from 5.0 percent to 2.4 percent. Sludge mass increased by about 91 kg (200 lb)/day. A new secondary digester was constructed to handle the additional volume. The volume and mass of digested sludge increased, raising disposal costs.	Digested sludge TS is 5.0 percent. Supernatant contains 0.2 percent TS.
Merrickville, OH	3,030 (0.8)	Iron-secondary from an extended aeration activated sludge plant without primary treatment	The plant began operation with phosphorus removal on-stream. Therefore, there is no comparison between digester operation with and without the iron sludge. However, the following operational problems resulted from design errors: (1) the air lift pumps could not remove the heavy sludge from the digester fast enough; (2) the mass of sludge was greater than anticipated, so the air supply rate was greater than predicted.	

(continued)

TABLE 14 (continued)

Location	Size-m^3/day (mgd)	Type of Sludge Treated	Impacts of Chemical Sludge	Performance with Chemical Sludge
Addison, MI	1,510 (0.4)	Alum-primary from an aerated lagoon plant	The sludge volume fed to the digester increased and the sludge was heavier and more difficult to pump. Mixing the sludge in the digesters was also more difficult so the air supply rate had to be raised to accomplish this.	The digester feed is 5.7 m^3/day (1,500 gal/day) primary sludge at 3.3 percent TS. The digested sludge is 6 percent TS. Supernatant concentrations are 320 mg/ℓ SS and 117 mg/ℓ BOD_5.
Columbia Boro, PN	4,920 (1.3)	Alum-secondary from a contact stabilization activated sludge plant without primary treatment	Sludge volume decreased from 113 to 90 m^3/day (30,000 to 24,000 gal/day)as TS increased from 0.5 percent to 1.25 percent. An additional 230 kg/day (500 lb/day) dry TS is fed to digester. Digester supernatant volume has decreased. The volume and mass of digested sludge has increased, raising disposal costs.	The digested sludge is 3 percent to 4 percent TS. Supernatant return is 68 m^3/day (18,000 gal/day) at less than 0.5 percent TS containing 500 to 1,000 mg/ℓ BOD_5.

was the most common disposal method used. The digested sludge TS concentrations varied from 3 to 6 percent.

No problems with deterioration of supernatant quality were reported as the result of either the additional sludge feed quantities or the changed sludge characteristics. This is significant because poor supernatant quality is often an indicator of digester overloading. Apparently, the capacity of the existing aerobic digesters was usually adequate to handle the additional quantities of chemical sludges. One plant reported the need for additional secondary digester capacity to handle the chemical sludge, but did not require additional primary digester capacity.

Literature

Accounts of aerobic digestion of chemical sludges available in the literature can be broken down into laboratory-scale studies on the one hand and pilot- or full-scale studies on the other hand. The results of laboratory scale studies can be summarized as follows:

- Aerobic digestion of waste activated sludge is not affected to an appreciable degree, nor is it inhibited by the presence of iron or aluminum precipitates (28).

- The release of soluble organic carbon and nutrients into the liquid phase is not enhanced by the presence of iron and aluminum precipitates (28).

- An aeration period of 10 to 15 days provides satisfactory stabilization of iron- and alum- waste-activated sludges at a temperature of 20°C (28).

- Dewatering characteristics of digested iron- and alum-waste activated sludges are poor, especially when long aeration periods are employed in batch treatment (28).

- With iron- and alum-waste-activated sludges, batch digester operation results in a greater destruction of sludge volatile solids and a lower sludge oxygen uptake rate than semi-continuous operation. However, the latter method provides better sludge dewaterability and better supernatant quality (28).

- Aerobic digestion can be successfully applied to lime-primary sludges, and the kinetics of the digestion process are not appreciably affected except at high lime dosages (33).

- When treating lime-primary sludges, the digestion system shows good buffering capacity, and the pH of the system is always maintained at 8.5 or above. The latter is contrary to the process characteristics with primary or waste-activated sludge alone, which are a low buffering capacity and a pH reaching as low as 4.0 (33).

- Aerobically digested lime-primary sludges have good settling and dewatering characteristics (33).

It should be kept in mind that the above conclusions were based on laboratory research. An experimental study of full-scale plant operation was conducted at the Portage Lake, Michigan, wastewater treatment plant (2). The plant influent flow of 6,400 m^3/day (1.7 mgd) was split after passing through the aerated grit chamber. The rest of the plant was divided into two identical halves, each side consisting of a contract stabilization activated sludge system and an aerobic sludge digester. The two 765-m^3 (201,990-gal) digesters were single-stage, batch-type operation units. During the study, one half of the plant was fed an alum dosage of 84 mg/ℓ just ahead of the contact basin, while the other half did not receive chemical feed.

Alum addition increased the aerobic digester sludge feed volume by 16 percent and the dry weight by 50 percent. It increased the digested sludge volume by 37 percent and the dry weight by 92 percent. However, the test was carried out in the summer, and it was projected that under average annual conditions there would be only a 20 percent increase in digested sludge volume and a 66 percent increase in dry weight. The alum-biological sludge was less amenable to aerobic digestion than the biological sludge. The alum-biological sludge TSS reduction observed through aerobic digestion under the summer conditions of the study was relatively low, on the order of 12 percent. It was predicted that the average annual rate would be even less (2).

The aerobically digested sludge gravity thickening capacity, as measured in laboratory cylinders, was higher for the alum-biological sludge than for the biological sludge. Batch laboratory-scale flotation thickening tests indicated no difference between the thickening properties of the two digested sludges. The alum sludge was concentrated by flotation thickening from 0.75 to 3.5 percent SS without polymer addition at an air to solids ratio of 0.03. Under these apparently optimal conditions, the underflow SS concentration was 88 mg/ℓ SS. It is likely that addition of cationic polymer could have improved the performance (2).

Laboratory vacuum filtration tests conducted by the Buchner funnel method indicated that the alum-biological sludge filtered slightly better than the biological sludge. Both sludges were conditioned with ferric chloride. Vacuum filter leaf tests indicated that the conditioned alum-biological sludge could be dewatered from 1.6 percent SS to a cake concentration of 16 percent SS at a filtration rate of 14.6 kg/m^2/hr (3.0 lb SS/hr/sq ft) (2).

An evaluation of the feasibility of sludge dewatering by centrifugation was conducted using a 305-mm (12-in) diameter pilot-scale basket centrifuge. Several runs were performed at various feed rates and feed solids concentrations. After application of the manufacturer's scale-up procedure, the results indicated that a full size 1.22-m (48-in) diameter basket centrifuge could dewater the digested alum-biological sludge at 1.6 percent feed SS to a solid cake containing 16 percent TS at a rate of 158.9 kg dry solids/hr (350 lb/hr) with no chemical addition. The SS concentration ranged from 300 mg/ℓ to 600 mg/ℓ, and the solids recovery was 96 percent (2).

Because of the good thickening characteristics of the alum-biological sludge, it was predicted that a more concentrated sludge could be removed from the digester by converting it to a two-stage unit. Plans were made to implement this by installing a concrete partition in the existing digester and an air lift pump for transferring sludge from the bottom of the first stage to the second stage. In year-round operation, it was estimated that digested alum-biological sludge of 1.25 percent SS would be obtained with the one-stage and 1.7 percent SS with the two-stage process (2).

Based on the study results, the costs of various alternatives for further sludge processing and disposal were considered. All of the alternatives assumed ultimate disposal on land at a location about 11 km (7 mi) from the plant. The alternatives were: direct trucking by tanker at 1.7 percent SS; centrifugation or vacuum filtration and hauling cake at 16 percent TS; and flotation thickening and tanker hauling at 4 percent SS. Table 15 summarizes the estimated costs of each alternative (2).

TABLE 15 PROJECTED COSTS OF VARIOUS SLUDGE HANDLING ALTERNATIVES FOLLOWING AEROBIC DIGESTION AT PORTAGE LAKE, MICHIGAN, $ (2)

	Direct Trucking	Basket Centrifuge	Vacuum Filter	Flotation
Capital investment	79,000	143,000	137,000	202,000
Annual debt service	8,951	16,203	15,523	22,888
Annual operation and maintenance	16,109	12,046	14,795	18,209
Total annual cost	25,060	28,249	30,318	41,097

The estimated capital costs included all mechanical equipment, building space, and vehicles required. Operation and maintenance costs included labor, power, fuel, chemicals, snow removal, and disposal site maintenance. Aerobic digestion costs were not included. Annual debt service was based on 7.5 percent interest over a useful life of 15 years. All costs were intended to reflect January 1975 levels. The projected costs showed direct trucking to be the most economical alternative primarily because of the relatively short haul distance. However, if the one-way haul distance were increased from 11 km to 16 km (10 mi), the centrifugation system would yield the minimum annual cost (2).

Conclusions

Aerobic digesters can be a cost-effective means of stabilizing chemical sludges when attempts are made by plant management to find procedures yielding maximum efficiency. Often, however, plant personnel prefer to operate them merely as aerated sludge holding tanks. The research conducted at Portage Lake, Michigan, (2) showed that the TSS reduction observed through aerobic digestion of alum-waste activated sludge was relatively low compared to that observed for waste activated sludge, but that the alum sludge thickened to a higher solids concentration. The investigators concluded that the chief utility of aerobic digestion of chemical-biological sludge may be sludge storage and thickening. No other reports of poor digestibility of chemical sludges were found in the literature.

The information which has been presented suggests the following ways of modifying aerobic digester operation to achieve better performance when treating chemical sludges:

- Converting a one-stage to a two-stage system. This is reported to increase the digested sludge solids concentration and sludge dewaterability, while not adversely affecting supernatant characteristics. This method may, however, yield less volatile solids destruction and require a higher oxygen supply rate.

- Installing a new secondary digester (or converting an unused existing basin) to handle the additional chemical sludge. It may not be necessary to install a new primary digester for moderate increases in sludge volume.

Although no reports of poor solids-liquid separation in aerobic digesters treating chemical sludges were found in the present study, there are situations in which this problem will occur. In this case, the addition of polymer may aid settling of the sludge in the digester and reduce the volume for disposal. This method could result in significant savings by reducing trucking costs for hauling liquid sludge. At Escanaba, Michigan (35), solids-liquid separation could not be achieved in aerobic digesters treating waste activated sludge. Addition of 25 to 30 mg/ℓ of a cationic polymer resulted in sludge concentration of 1.8 to 2.2 percent TS.

COMPOSTING

Composting is being viewed with increasing interest as a sludge management alternative. It is appealing because it converts sludge into a product that is aesthetically acceptable, essentially free of pathogens, and easy to handle. The compost produced can be used to improve soil structure, increase its retention of water, and provide nutrients for plant growth. However, the fact that the process may be relatively expensive (compared to land application of liquid digested sludge, for example) has prevented its widespread use in the past. Several investigations in the United States and Canada are presently underway to determine if composting can be made economical for widespread use.

Since sludge characteristics are an important factor in treatment costs, the economics of composting are likely to differ for chemical sludges and regular sludges. In the preceding sections of this report, the characteristics of regular sludges and chemical sludges resulting from phosphorus removal were compared. In this section, information related to the effects of phosphorus removal on sludge composting is presented. It is intended to help answer questions about the ability of chemical sludges to dry as expected to remain aerobic, and to support biological growth. It will also look at the value of the chemical sludge compost as a fertilizer and soil conditioner.

Questionnaire Results and Case Studies

As previously shown in Table 10, only two (1 percent) of the plants responding to the questionnaire survey were composting their chemical sludges. Both of these plants were case study sites. At the Midland, Michigan, site, an informal sludge composting operation was underway. The operation utilized vacuum filtered sludge cake containing iron precipitates from the plant's primary addition of iron for phosphorus removal. The sludge cake may have been particularly suitable for composting because of the plant's thermal conditioning step. Thermal conditioning enabled production of a filter cake which was very dry (about 50 percent dry TS); it pasteurized the sludge; and it eliminated the need for using chemical conditioners such as iron or lime which would further increase the chemical content of the sludge cake.

All of the plant's 4.6 m^3/day (10.8 yd^3/day) of filter cake was transported approximately 6.5 mi to the city of Midland's sanitary landfill site, but not all of it was composted. Some of the sludge was used for land reclamation at the landfill site. The remainder was informally made into compost by the city's

Department of Forestry and used as soil conditioner in their onsite ornamental tree nursery. Both sludge-leaf compost and sludge-sawdust compost were made in piles, with a minimum of attention and labor from the City Forester and a few employees. The compost piles were simply turned over once every three days.

The compost greatly enriched and conditioned the native hard clay soil at the site. The texture of the soil was greatly improved by mixing in compost. Over part of the tree nursery, 15 to 24 cm (6 to 8 in) of sludge cake was disced directly into the soil, while on other plots, up to 0.6 cm (2 ft) of sludge-leaf compost was used. The sludge-leaf compost was not as rich a fertilizer as the sludge disced in alone, but it may have improved the texture of the soil more. Whether sludge-leaf compost or sludge alone was disced in, placing a sludge-sawdust mixture (1/2 sludge - 1/2 sawdust) on top of the soil produced even richer growth. Iron sludge composting at Midland, therefore, was both successful and valuable to the city.

It has not been possible to assess the costs involved because the operation was very informal and the quantity of sludge used was small. It can be reasoned, however, that the costs would have been raised as the result of iron addition for phosphorus removal had not the initial impact of iron addition on filter cake moisture content been overcome. The initial impact was an increase in filter cake moisture content, meaning higher costs for sludge cake hauling and perhaps a slower rate of composting. Filter cake dryness was restored, however, by raising the temperature of the thermal conditioning unit, thus avoiding these problems at Midland.

At Windsor, Ontario, a sludge disposal arrangement existed between the Little River Pollution Control Plant and a commercial composting operation. After the plant hauled the sludge 11 mi to the composting site, it payed \$2.20/t (\$2.00/ton) to the commercial operation to handle the sludge from there. The commercial firm mixed the sludge with sawdust where it is dumped from the truck in the field. The sludge then dried in the field without being turned over. When dry, it was taken into a barn where it was worked considerably and piled. The temperature inside the piles reached 66° to 71°C (150° to 160°F). The operation was monitored by the Ontario Ministry of the Environment to see that it met health standards. The product was shredded and then bagged. Further details of the private operation were proprietary. The sludge-sawdust compost material was sold in bags for \$198/t (\$180/ton) retail and \$88/t (\$80/ton) wholesale. Unbagged compost was sold in bulk for \$16/m^3 (\$12/yd^3). The product was marketed as a soil additive and organic fertilizer primarily for home gardening and commercial greenhouse applications.

At Little River, alum was added to the raw sewage for phosphorus removal. The undigested sludge was centrifuged before being trucked to the composting site. Alum addition resulted in an increase in the average moisture content of the centrifuge cake from 80 percent to 84 percent. This increase in moisture content raised the costs of hauling the sludge cake, and probably also slowed the rate of drying of the sludge in the compost piles. The owner of the composting operation reported having no problems as the result of the presence of aluminum precipitates in the sludge. However, he felt that a lime sludge would make a better product, which would cost perhaps $2.20/t ($2.00/ton) less to make.

Presently, the cost to the plant to dispose of 9,604 t/yr (10,589 tons/yr) of dry sludge TS from the centrifuges is estimated at $3.73/t ($3.38/ton). This includes the cost of sludge hauling and the fee paid to the composting operation to take the sludge.

The City of Windsor was planning to start a composting operation at its other plant, the West Windsor Pollution Control Plant. Sludge was to be composted in static piles with forced ventilation. Wood chips were to be used as the bulking agent. Preliminary tests indicated that there was no difficulty in composting the plant's undigested iron-primary sludge. The cost of composting 31,750 t/yr (35,000 tons/yr) of dry sludge TS was estimated at $7.50/t ($6.80/ton) of dry solids, a very low estimate compared to many others found in the literature for sewage sludges. The estimate included no land capital cost because the land was already owned by the city. Similarly, there were to be no sludge hauling fees since the land was adjacent to the plant. A total site preparation cost of $85,000 was estimated, with an equipment cost of $114,000. Assuming a ten-year life span for the equipment and site preparation, the total capital cost was $20,000/yr, or $0.66/t ($0.60/ton).

The operating costs were estimated at $217,000/yr or $6.84/t ($6.20/ton) dry sludge. The value of the compost produced was expected to be about $8.82/t ($8.00/ton) in bulk sales, making the project very economical.

Conclusions

At at least two sites in the United States and Canada, a valuable compost was produced from all or part of the chemical sludge from a municipal wastewater plant with phosphorus removal. Based on the experiences of these plants, it appears that the presence of iron and aluminum precipitates in the sludge does not inhibit the growth of the microorganisms which are necessary for composting. The quality and value of the compst as a soil conditioner and fertilizer can be very high for iron and aluminum sludges. It is likely that the value of a lime sludge compost would be even higher because of the beneficial effects of the lime on soil pH; and one compost manufacturer thought that lime sludge compost would be cheaper to produce by the ton.

There may be some problems with composting of iron and aluminum sludges, however. The chemical sludges often cannot be dewatered by vacuum filter or centrifuge to a moisture content as low as that to which regular sludges can be dewatered. When this is the case, the cost of hauling the sludge cake to the composting site increases. In addition, a wetter sludge could be expected to require more time for drying and composting to occur, increasing land requirements and operating costs. More bulking agent and/or more ventilation of the compost piles could also be necessary to keep the wetter sludge aerobic.

SECTION 7

CONDITIONING OF CHEMICAL SLUDGES

CHEMICAL CONDITIONING

Among the plants participating in this study, 40 (23 percent) out of 174 practiced sludge conditioning with chemicals such as lime, ferric chloride, and polyelectrolytes. The survey showed chemical conditioning to be much more than other conditioning methods for chemical sludges.

The purpose of chemical conditioning is to improve the performance of subsequent dewatering processes. Maximum dewatering performance requires the optimization of many more operating variables for chemical conditioning than for thermal conditioning. The type(s) of chemical conditioner(s), the dosage(s), and the method(s) of application must be chosen, with a wide range of choices and combinations possible. The success of the conditioning process can only be measured by its effects on subsequent dewatering, incineration, and disposal processes. Therefore, chemical conditioning has been discussed in each of the sections of this report dealing with dewatering and also in the section on incineration. The reader is referred to those sections for detailed information from the literature and from the field investigations and questionnaire survey which were part of this study.

In general, the most common impact of phosphorus removal on chemical conditioning requirements for sludges has been the raising of the dosages required for successful dewatering. Increased chemical dosages are undesirable not only because of the effect on incinerator auxiliary fuel requirements: with inorganic and inert chemical conditioners, increased chemical dosages provide a drier cake on a total solids basis, but the cake is wetter on a sludge (as opposed to total) solids basis. That is, the kg water/kg dry sludge solids fed to the incinerator is higher. Thus, more fuel is required for incineration.

As far as auxiliary fuel requirements, the polyelectrolyte conditioners have an inherent advantage, and the inert materials such as fly ash and sludge ash have inherent disadvantages. For the same cake TS content, the polyelectrolyte-conditioned sludge cake will require significantly less auxiliary fuel while the ash-conditioned sludge cake significantly more auxiliary fuel than the lime and ferric chloride-conditioned cake.

With regard to chemical sludges, then, particular attention should be paid to cases where polyelectrolyte conditioners have been successful in producing dry sludge cakes and have kept down incinerator fuel requirements. Also promising are cases where, because of the presence of phosphorus removal chemicals in sludge, fewer chemicals were required for conditioning. In scanning the sections of this report on dewatering, the reader will note that in few instances have these ideal situations occurred, but in some cases such as pressure filtration of an iron sludge and centrifugation of lime sludge, there were promising results.

THERMAL CONDITIONING

Introduction

Thermal conditioning is a technique which has been increasingly used by consulting engineers in recent years. Two types of thermal conditioning exist. The Zimpro process, often termed low-pressure oxidation, differs from the Porteous or Farrer processes in that air is injected into, rather than excluded from, the process reactor. Thermal sludge conditioning has several benefits. It stabilizes the sludge, enabling further handling without pathogens, odors, or putrefaction. It changes the cellular structure of the sludge, enabling thickening to a relatively high solids concentration and thus reducing the volume to be vacuum filtered; and it improves the dewatering characteristics of the sludge, increasing filter yield and cake solids concentration. Its disadvantages are the capital and operating costs of the process and production of a high COD decantate.

Questionnaire Survey

As previously shown in Table 10, 11 (6 percent) of the plants responding to the questionnaire survey reported that they have thermal conditioning. Of these 11 plants, 7 practiced anaerobic digestion before the thermal conditioning step. The digested and thermally conditioned sludge from these plants was either applied to drying beds, applied directly to croplands, or vacuum filtered. At the other four plants which did not anaerobically digest, a gravity thickening step preceded thermal conditioning. At these plants the thermally conditioned sludge was either vacuum filtered or centrifuged.

Only three of the questionnaire respondents commented on the performance of thermal conditioning at their plant or the impact of the chemical sludge. At the Maumee River plant in Waterville, Ohio, a contact stabilization activated sludge plant, a combined alum-secondary and primary sludge was treated. The sludge was gravity thickened, underwent Zimpro, Inc. low-pressure oxidation, and was dewatered on a belt vacuum filter. The plant used polymer for further conditioning of the sludge before dewatering. A filter cake of 35 percent TS was achieved.

At the Grand Haven, Michigan, plant, a 14,000 m^3/day (3.7 mgd) modified activated sludge plant, a combined iron-primary and waste activated sludge was treated. The raw sludge had an average TS concentration of 4.38 percent. The VS fraction was 65 percent of TS. The sludge was gravity thickened, underwent Zimpro, Inc. low-pressure oxidation, and was vacuum filtered and incinerated. The thermally conditioned sludge had a TS concentration of 11.8 percent. Although the sludge was very difficult to dewater because of a tannery waste received by the plant, thermal conditioning was enabling a filter cake TS concentration of over 30 percent. The high TS concentration of the filter cake enabled economical operation of the incinerator. A comparison of the characteristics of Grand Haven's filtrate, thickener supernatant, and thermal conditioning decantate is shown in Table 16.

The plant was looking for an alternative to thermal conditioning for use when the unit was down for repair. Chemical conditioning had been tried with more than 50 different polymers, but none of them produced a filter cake of greater than 19 percent TS.

The North Olmsted, Ohio, plant, a contact stabilization activated sludge plant, treated combined aluminum-secondary and primary sludge. The sludge was gravity thickened, underwent Zimpro, Inc. low-pressure oxidation, and was vacuum filtered. The plant achieved a filter cake of 46 percent TS and reported that the performance was not affected by the addition of sodium aluminate for phosphorus removal. The thermal conditioner decantate contained 3,000 mg/ℓ SS and 2,631 mg/ℓ BOD.

Case Studies

The case studies in the appendices contain detailed information on the effects of chemical sludges on thermal conditioning. The particular case studies which deal with this subject are similar to the experiences of other plants responding to the questionnaire survey. However, unusual highlights of the case studies can be mentioned.

At Port Huron, Michigan, the costs of thermal conditioning using the Farrer system were compared with the costs of chemical conditioning with polymer. The plant practiced centrifugation and incineration of the conditioned sludge. When thermal conditioning, the plant reduced its costs for conditioning polymer and fuel oil (for incineration) by \$12.63/t (\$11.46/ton) and \$37.71 t (\$34.20/ton), respectively. The operational and maintenance expenses involved in thermal conditioning included \$20.94/t (\$19.00/ton) for natural gas, \$1.08/t (\$0.98/ton) for electricity, \$1.76/t (\$1.60/ton) for boiler water conditioning and cleaning chemicals, \$1.68/t (\$1.52/ton) odor control, and \$9.52/t (\$8.64/ton) for sidestream treatment. The major operational and maintenance concern, however, was the cost of supplies and labor

TABLE 16. AVERAGE SIDESTREAM CHARACTERISTICS AT GRAND HAVEN, MICHIGAN

Type of Sidestream	Thickener Supernatant	Thermal Conditioner Decantate	Filtrate
Volume (m^3/day (gal/day))	889 (235,000)	140 (37,000)	31 (8,230)
SS (mg/ℓ)	425	2,295	2,358
VSS (% of SS)	70.3	60.8	62.6
COD (mg/ℓ)	N/A	20,900	18,267
BOD (mg/ℓ)	173	7,737	6,755

for equipment maintenance and repair. This cost was unknown, because the unit was in use for a limited time only. The equipment suffered from extreme corrosion and erosion of heat exchanger and connecting piping, and its operation was eventually discontinued to alleviate the need to replace all the parts with stainless steel. Polymer conditioning was then relied upon. The case concluded that if the unit could have been operated at an equipment maintenance and repair cost of less than $15.35/t ($13.92/ton), it would have been more cost-effective than polymer conditioning.

Without thermal conditioning, the sludge fed to the centrifuges was of a lower solids concentration, and more time was needed to dewater it. The filter cake solids concentration was also reduced, and the capacity of the incinerator was reduced by about 454 kg/hr (1,000 lb/hr). Because of the high costs of operating under these conditions, the plant was evaluating several alternative treatment systems.

At Midland, Michigan, ferric chloride addition was found to have adversely affected the performance of the Zimpro, Inc. thermal conditioner. The impaired performance was evidenced mainly by the poorer vacuum filter yield and lower filter cake TS concentration. It was found that excellent performance could be restored by raising the temperature of the thermal conditioner from 185°C (365°F) to 202°C (395°F).

The effect of ferric chloride addition on the thermal conditioner decantate SS and BOD concentrations was unknown. During ferric addition, the decantate was low in SS (478 mg/ℓ) but high in BOD (6,000 mg/ℓ). The vacuum filter filtrate SS concentration was high before, and even higher after, ferric addition began. The BOD concentration of the filtrate was also high before ferric addition began but was not further increased by ferric addition.

When alum addition for phosphorus removal was tried at Midland there were poor conditioning and dewatering results even at the higher thermal conditioner temperature.

Literature

A single literature account of thermal conditioning of chemical sludges was found (26). Dewatering of the thermally conditioned sludges by vacuum filtration, centrifugation, or on drying beds was studied. Three plants using the Zimpro, Inc. low-pressure oxidation system were involved. The characteristics of these plants are described in Table 17. At the Midland and Lucas County plants, the thermally conditioned sludges were vacuum filtered, while at Defiance sludge was applied to drying beds. Some of the Defiance sludge was anaerobically digested rather than thermally conditioned before being applied to the drying beds. Data from the plants were gathered which characterized the

TABLE 17. CHARACTERISTICS OF PLANTS IN STUDY OF THERMAL CONDITIONING OF CHEMICAL SLUDGES (26)

Plant Location	Type of Raw Sludge Treated (m^3/day (mgd))	Reactor Temperature (^{o}C (^{o}F))	Reactor Pressure (kg/cm^2 (psi))	Further Treatment/ Disposal
Defiance, Ohio	Iron-primary and secondary from an activated sludge plant	171 (340)	16.52 (235)	Sludge drying beds; dried sludge stock-piling on land adjacent to plant
Midland, Michigan	Iron-primary and secondary from a trickling filter plant	193 (380)	31.63 (450)	Vacuum filters; land disposal of filter cake
Lucas County, Ohio	Aluminum-primary from a contact stabilization activated sludge plant with no primary treatment	182 (360)	21.09 (300)	Vacuum filters; land disposal of filter cake

sidestreams associated with thermal conditioning, vacuum filtration, and anaerobic digestion, and are presented in Table 18. In the case of the decantate from the three thermal conditioning units, the TS concentrations averaged 1 percent, with a VS fraction of 74 to 87 percent of TS. The decantates contained 0.3 to 0.6 percent organic carbon, almost all of which was soluble in each case. These observations indicate that the decantate exerted a high oxygen demand on the plant biological oxidation processes when it was returned to the plant. Sidestreams from subsequent dewatering processes would be expected also to have been high in soluble organic carbon. The amount of phosphorus, and either iron or aluminum, in the thermal conditioner decantate varied widely between the plants.

The Midland thermally conditioned iron sludge was dewatered in a pilot-scale, solid-bowl scroll centrifuge at two different bowl speeds and loading rates. The results of this test are given in Table 19. At the higher of the two bowl speeds and loading rates tested, both the centrate TS concentration and the percentage of feed solids recovered in the cake were greater than at their lower rates.

Table 20 contains a comparison of the sidestream characteristics from plant vacuum filter operations and pilot-scale centrifugation of the plant sludges. In the two comparisons made, the vacuum filter filtrate was lower in organic carbon, phosphorus, and iron or aluminum than the scroll or basket centrifuge centrate.

Conclusions

Few problems have been experienced in thermal conditioner operation as the result of chemical addition for phosphorus removal. Methods of improving thermal conditioner operation when treating chemical sludges may include:

- Providing sludge storage facilities or alternate disposal methods such as liquid sludge hauling for periods when the thermal conditioner is down for repair. This is particularly important at plants practicing incineration because chemical conditioning as a substitute may not provide a filter cake of adequate dryness. Repair requirements for thermal conditioners seem to be moderate to substantial.

- Raising the temperature of the unit. This may reduce decantate SS concentrations and improve sludge dewaterability.

There is some indication that vacuum filters may be more suitable for the dewatering of thermally conditioned chemical sludges than centrifuges. However, this should be investigated further, and filter presses should also be evaluated.

TABLE 18. CHARACTERISTICS OF STREAMS AND SIDESTREAMS ASSOCIATED WITH SLUDGE TREATMENT OPERATIONS (26)

Stream or Sidestream	Fe or Al mg/ℓ	P mg/ℓ	Fe or Al to P ratio	TOC mg/ℓ	SOC mg/ℓ	TS mg/ℓ	% VS
Lucas County Raw sludge	2,200	1,157	2.0	4,800	2,441	29,450	57
Midland Raw sludge	7,250	1,867	3.9	41,830	5,467	86,333	54
Lucas County Conditioned sludge	3,045	1,955	1.5	10,650	4,950	47,412	53
Midland Conditioned sludge	12,233	4,317	2.8	--	--	200,000	41
Defiance Conditioned sludge	5,570	1,483	3.3	52,400	5,019	110,000	48
Lucas County Decantate	5	20	0.2	2,937	2,887	9,183	87
Midland Decantate	70	723	0.1	6,000	5,800	10,667	76
Defiance Decantate	122	66	1.8	5,090	3,902	10,340	74
Defiance Digested sludge	2,545	656	3.9	10,600	2,050	35,000	47
Defiance Supernatant	635	26	2.4	6,400	2,180	17,500	53
Midland Filter cake	73,000	17,000	4.2	500,000	7,400	598,000	42
Midland Filtrate	5,200	2,300	2.3	4,000	3,200	7,200	62

(continued)

TABLE 18 (continued)

Stream or Sidestream	Fe or Al mg/ℓ	P mg/ℓ	Fe or Al to P ratio	TOC mg/ℓ	SOC mg/ℓ	TS mg/ℓ	% VS
Lucas County Filter cake	30,336	15,218	2.0	--	--	338,000	45
Lucas County Filtrate	95	69	1.4	3,811	3,489	7,878	82
Lucas County Filter cloth wash	100	64	1.6	4,700	4,500	8,500	85
Lucas County Digested sludge	1,680	950	1.8	2,100	270	18,300	43

TABLE 19. RESULTS OF PILOT SCALE CENTRIFUGATION OF MIDLAND THERMALLY CONDITIONED IRON SLUDGE (26)

Bowl Speed, rpm	Loading Rate kg/hr (lb/hr)	Cake Solids % TS	Centrate Solids % TS	Polymer	Solids Recovery, %
4,000	19.2 (42.3)	33.3	4.5	no	82.9
4,500	24.7 (54.4)	49.0	9.6	no	97.1

TABLE 20. COMPARISON OF SIDESTREAMS FROM PLANT AND PILOT DEWATERING OPERATIONS (26)

Operation	Fe or Al to P ratio	P mg/ℓ	TOC mg/ℓ	SOC mg/ℓ	TS mg/ℓ	% VS
Midland:						
Vacuum filtration filtrate	520	230	4,000	3,200	7,200	62
Scroll centrifugation centrate	640	170	5,600	4,400	9,600	64
Lucas County:						
Vacuum filtration filtrate	95	69	3,811	3,489	7,878	82
Scroll centrifugation centrate	458	354	6,560	4,180	14,500	68
Basket centrifugation centrate	260	200	9,800	2,000	8,000	84

SECTION 8

DEWATERING OF CHEMICAL SLUDGES

DRYING BEDS

Introduction

A total of 58 (33 percent) of the plants responding to the questionnaire survey utilized drying beds for sludge dewatering. Historically, sludge drying beds have been mainly used by smaller communities. Principal advantages of drying beds are their simplicity and low maintenance costs. The chemical cost and operating complexity of mechanical dewatering equipment are additional factors which favor drying beds. Disadvantages include their large land requirements, inability to dewater effectively year-round in certain areas, and potential odor problems.

Questionnaire Survey

Nine of the questionnaire respondents which have drying beds reported problems with the handling of chemical sludges. All of those plants used iron or aluminum salts for phosphorus removal. Table 21 summarizes the experiences of those plants. Plants using lime did not report any impacts, either positive or negative, on drying bed operation other than the need for additional bed space to handle the extra sludge generated.

The most common problems experienced in drying bed operation were increases in sludge volume and mass, and poor sludge dewaterability. These problems resulted in: construction of additional drying beds, addition of chemicals to improve dewatering; modifications in sludge application procedures, lengthened sludge drying times, or the abandonment of sludge drying beds in favor of another dewatering process. The point of chemical addition did not seem to influence the performance of iron or aluminum sludges on drying beds. Plants with chemical addition to primary or secondary stages reported the same general problems.

Parry Sound, Ontario, a primary plant adding ferric chloride for phosphorus removal, reported rapid clogging of their sand drying beds with solids. This problem was alleviated by replacing the original bed sand with a coarser sand and by replacing the bed sand more frequently.

TABLE 21. IMPACTS OF CHEMICAL SLUDGES UPON DRYING BED PERFORMANCE AS REPORTED IN QUESTIONNAIRE RESPONSE

Location	Size-m^3/day (mgd)	Type of Sludge Treated	Impacts of Chemical Sludge	Performance with Chemical Sludge
Addison, Michigan	1,480 (0.39)	Alum-primary (aerobically digested) from an aerated lagoon plant.	The chemical sludge takes longer to dewater on the beds.	
Escanaba, Michigan	7,040 (1.86)	Iron-primary (anaerobically digested) from an activated sludge plant.	The sludge is not more difficult to dewater, but there is a larger volume. A method was devised to add polymer to the sludge as it enters the drying beds to increase the bed turnover rate. Also, additional cement strips were added in the drying beds to allow mechanical sludge removal. This also increases the bed turnover rate.	
Rogers City, Michigan	2,650 (0.7)	Iron-secondary (anaerobically digested) from a conventional activated sludge plant with no primary treatment.		The sludge at 3.2 percent TS is applied to sand drying beds. When applied 8 to 12 in deep, the sludge is ready for removal by hand in 30 to 45 days. It is best to fill the beds to only

(continued)

TABLE 21 (continued)

Location	Size-m^3/day (mgd)	Type of Sludge Treated	Impacts of Chemical Sludge	Performance with Chemical Sludge
Rogers City, Michigan (cont'd)				8 in for fastest drying and easiest removal.
Flushing, Michigan	4,580 (1.21)	Iron-secondary and primary (anaerobically digested) from a conventional activated sludge plant.		The estimated operation and maintenance costs for the drying beds were about $700 in 1976.
Charlevoix, Michigan	1,100 (0.29)	Iron-secondary and primary (anaerobically digested) from a complete mix activated sludge plant.	Before phosphorus removal was begun, the plant had only primary sludge. The sludge at that time took one half as long to dewater and was slightly less odorous.	
Coldwater, Michigan	6,780 (1.79)	Iron-primary and secondary (anaerobically digested) from a trickling filter plant.	The sludge volume was three times greater after phosphorus removal, causing the plant to construct 7 new drying beds. Cement bottoms were put in both the old and new beds. The sludge takes 30 to 50 percent longer to dry.	

(continued)

TABLE 21 (continued)

Location	Size-m^3/day (mgd)	Type of Sludge Treated	Impacts of Chemical Sludge	Performance with Chemical Sludge
Parry Sound, Ontario	3,200 (0.85)	Iron-primary (anaerobically digested) from a plant with no secondary treatment.	The sludge is finer and more difficult to dewater. To improve dewatering, a more coarsely screened sand was put in and the sand is changed more often.	
Ashland, Wisconsin	4,540 (1.2)	Alum-secondary and primary (anaerobically digested) from a conventional activated sludge plant.	Sludge drying beds were abandoned because the gelatinous alum sludge would not dewater. The plant began using a vacuum filter instead.	
Three Rivers, Michigan	4,730 (1.25)	Alum-primary and secondary (anaerobically digested) from a conventional activated sludge plant.	The sludge takes longer to dry on the beds.	

The drying beds at Ashland, Wisconsin, an activated sludge plant with secondary addition of aluminum sulfate, were abandoned because of difficulties in dewatering the gelatinous alum sludge. This plant switched to ferric chloride sludge conditioning and vacuum filter dewatering.

Literature

Literature accounts of drying bed dewatering of chemical sludges are few, although the recent rise of chemical addition for phosphorus removal should change this pattern.

An excellent discussion of chemical sludge dewatering on sand beds (48) relates sludge characteristics to sand bed design. It concludes:

- The most critical sludge characteristic with regard to the use of sand beds is sufficient sludge compressibility to prevent sand bed penetration. Compressibility is the variation of specific resistance with pressure. It influenced the rate of water loss of sludges applied to sand beds.
- Chemical sludge drainage rates depend upon the specific resistance and applied solids concentration.
- Air drying of chemical sludges occurs in two distinct phases. Initially, a slow drying rate occurs, followed by a more rapid drying.
- The slow drying rate appears to be governed by the applied depth and drained solids concentration. Rapid drying is approximately equal to the rate of free surface water evaporation.

The addition of a high molecular weight anionic polymer to the drying beds at the Escanaba, Michigan, plant had a dramatic effect on the sludge disposal operation (35). The plant's aerobically digested sludge failed to dewater and dry out in a reasonable length of time. The sludge had a tendency to penetrate and plug the sand bed. Polymer addition to the sludge as it entered the beds resulted in faster drying time. Before this operation, extensive hauling of liquid sludge for land disposal was necessary. The sludge was flocculated by placing a simple plywood mixing box (4 ft by 4 ft by 1.5 ft) under the bed discharge valve. The box contained a baffle to provide mixing and a notched overflow weir. Polymer was added to the sludge as it entered the box by means of a portable tank and a small electric pump mounted on a trailer. A used fuel tank (1.04 m^3 or 275 gal) with the top cut out made an excellent low cost tank. As the polymer was added to the mixing box, the sludge rate was adjusted until a large floc formed. The final concentration of polymer was dependent on the sludge TS concentration (Table 22). Several polymers were tried, with a high molecular weight anionic proving to be the most effective.

TABLE 22. POLYMER APPLICATION TO DRYING BEDS

Sludge % Solids	Cationic mg/ℓ	Anionic mg/ℓ
1.8	83	23
2.0	92	26
2.2	104	29

Drying time was evaluated by filling each of two beds with 56.8 m^3 (15,000 gal) of the same sludge. One bed had polymer added while the other did not. During the test period there was almost constant daily sunshine and minimal rainfall. At the end of 12 days, the bed with the polymer was dry enough to clean, while the one without polymer required 34 days to dry. During the summer, drying time varied depending on climatic conditions and the mass of dry sludge applied to the bed. Average drying time in ideal weather was 10 to 14 days with the polymer added.

Another benefit derived from the polymer addition was a reduction in odor from the drying beds. The relatively high volatile content of the digested sludge generated an offensive odor when it remained in the liquid state for long periods.

Conclusions

Improved drying bed performance can be achieved if the following modifications are made:

- Improving the performance of upstream facilities, e.g., thickeners, digesters
- Adding chemicals to improve sludge dewatering characteristics
- Optimizing sludge loading rates and bed turnover rates
- Changing the drying bed filter material
- Covering open beds where climatic conditions adversely affect performance.

VACUUM FILTRATION

Introduction

As shown in Table 10, vacuum filtration is utilized by approximately 21 percent of the sewage treatment plants responding to the project questionnaire. Vacuum filtration is an important sludge treatment unit process for many plants generating phosphorus-laden chemical sludges. It has received substantial attention in both the technical literature reviewed and the case studies conducted. It was reported that vacuum filtration operational characteristics and performance results were definitely affected by the inclusion of chemical sludges with primary and/or secondary sludges normally produced. The resulting changes were not in all cases detrimental, however, and solutions to the problems encountered were often achieved by treatment plant personnel.

In the following discussion it is assumed that the reader is familiar with the theory and mechanical operation of vacuum filters. The operator of a vacuum filter strives for maximum solids capture, filter cake yield, and filter cake solids content, as well as to minimize costs. Solids capture is usually expressed as the percent of the dry weight of sludge TS retained on the filter, i.e., that portion not returning to the treatment plant in the filtrate. A solids capture of at least 90 percent is desired.

Filter cake yield is expressed in terms of pounds of dry total sludge TS discharged from the filter media per hour per square foot of filter media. When chemicals are used for phosphate removal, they increase the percentage of non-volatile solids in the sludge. When comparing filter cake yields, correction should be made to account for this difference.

Cake solids content is expressed in terms of percent weight of dry TS in the sludge cake. The cake solids content normally increases with an increase in the TS concentration of the sludge being fed to the filter.

Questionnaire Survey

Table 23 summarizes the comments of 11 plants utilizing vacuum filtration which responded to the questionnaire survey. Many plants reported that vacuum filtration problems resulted from the inclusion of phosphorus-laden chemical sludges with the biological sludges previously processed. Typically, the increased sludge volume and solids mass stressed the capacity of the vacuum filter. In one extreme case, a plant reported increasing their vacuum filtration operation from 40 hr/wk up to 168 hr/wk. Doubling of operational time, e.g., from one to two shifts, was fairly common.

TABLE 23. CHANGES IN VACUUM FILTER PERFORMANCE REPORTED AS A RESULT OF PHOSPHORUS REMOVAL CHEMICAL SLUDGE ADDITION

Location	Type of Sludge	Changes in Vacuum Filter Performance Resulting from Phosphorus Removal Chemical Sludge Addition
London, Ontario	Lime-primary and secondary (conditioned with $FeCl_3$) from an activated sludge plant	Filter cake solids increased from 16.8 percent to 19.3 percent TS. Filter yield increased by 30 percent. Cost of chemical conditioning increased from $15 to $16 per ton of dry solids.
Sheboygan, Wisconsin	Iron-secondary and primary from a trickling filter plant	Filter yield decreased by 25 percent.
Hatfield Township, Pennsylvania	Lime-primary and alum-secondary	Tremendous increase in sludge volume, required going to a 24 hr/day, 7 day/wk operation on vacuum filter as opposed to the former 40 hr/wk operation.
Willoughby-Eastlake, Ohio	Alum-secondary and primary from an activated sludge plant	The sludge filter cake averages about 20 percent higher moisture content than formerly because of the alum addition. Filter discharge characteristics (solids capture) deteriorated. Experiments with various filter cloths showed that a sateen weave provided the best results. Lime and ferric chloride sludge conditioners also impared filter operation.

(continued)

TABLE 23 (continued)

Location	Type of Sludge	Changes in Vacuum Filter Performance Resulting from Phosphorus Removal Chemical Sludge Addition
Warren, Michigan	Alum-secondary and primary from an activated sludge plant.	Many other plant changes were incorporated concurrently with the implementation of phosphorus removal, making it impossible to identify specific changes caused by the alum addition. However, it appeared that the alum addition caused the sludge cake solids content to drop from 20 percent down to 16 percent. In addition, the sludge conditioning chemical was changed to polymer instead of lime and ferric chloride. As the result of more moisture in the sludge cake, subsequent incineration energy (natural gas) costs increased.
Ypsilanti, Michigan	Iron-secondary and primary from an activated sludge plant	The plant found it very difficult to handle the 50 percent increase in sludge generated by implementation of phosphorus removal. The vacuum filter operation time was increased by 50 percent, and solids increased in the filtrate and throughout the plant.
Lakeville, New York	Iron-tertiary and primary and secondary from a trickling filter plant	Changed from vacuum filtration of raw sludge to vacuum filtration of anaerobically digested sludge. More polymer was needed to treat the sludge.
Hilton, New York	Alum-secondary and primary from an activated sludge plant	This plant operates both vacuum filters and centrifuges. They found no changes due to alum addition for phosphorus removal. The vacuum filter yielded a 20 percent solids cake (sludge conditioned with lime and ferric chloride). The centrifuge yielded a 19 percent solids cake (polymer added).

(continued)

TABLE 23 (continued)

Location	Type of Sludge	Changes in Vacuum Filter Performance Resulting from Phosphorus Removal Chemical Sludge Addition
Lakewood, Ohio	Alum-secondary and primary from an activated sludge plant	The mixed liquor suspended solids concentration increased from 2000 to 5000 mg/ℓ, and the return sludge solids concentration increased from 1.2 percent to 2 percent. The vacuum filter operating time has increased to two shifts instead of one. The filter cake solids concentration has decreased to 20 percent from the former 25 percent because the alum sludge is more difficult to dewater. Overall the cost of plant operation has increased about 40 percent due to the addition of alum for phosphorus removal. The increased cost includes labor (3 men at $32,000/yr), lime and ferric chloride sludge conditioning at about $21,000/yr, furnace fuel at $83,000/yr, and increased hauling costs to the landfill.
Frankenmuth, Michigan	Alum-secondary and primary sludge (anaerobically digested) from an activated sludge plant	The luxury uptake of phosphorus in the activated sludge unit due to high BOD in the influent (brewery waste) has made the required phosphorus removal obtainable with a minimum amount of chemical addition. No problems have been experienced with vacuum filtration.
Milford, Michigan	Iron-secondary and primary from an activated sludge plant	The plant experienced greatly improved vacuum filter operation after the implementation of phosphorus removal. The vacuum filter yield increased from 24 kg/m^2/hr (5 lb/ft/hr) to 68 kg/m^2/hr (14 lb/ft/hr) and cake solids increased from 16 percent to 19 percent. In addition, there was a decrease in the quantity of lime and ferric chloride needed for sludge conditioning.

Also commonly reported was the requirement to change dosages or types of conditioning chemicals used (polymers, ferric chloride, etc.), as a result of different sludge characteristics caused by the introduction of chemical sludges.

Literature

Most literature sources indicated that increased filter cake yield can be expected from phosphorus-laden chemical sludges. Typical comments include the following:

- Laboratory vacuum filtration tests conducted by the Büchner funnel method indicated that the alum digester sludge filtered slightly better than the control sludge. Both sludges were effectively conditioned by ferric chloride at 4 percent by weight of TSS. Specific resistances were 10.9×10^5 sec^2/g and 7.6×10^5 sec^2/g for the control and alum digester sludges, respectively, after conditioning. Vacuum filter leaf tests indicated that the conditioned alum-biological sludge could be dewatered from 1.6 percent TSS at a filtration rate of 14.6 kg TSS/m^2/hr (3.0 lb/ft^2/hr) (2).

- A dual polyelectrolyte sludge conditioning sequence made it possible to attain vacuum filter yields of up to 49 kg/m^2/hr (10 lbs/ft^2/hr) on the phosphorus removal sludge solids (22).

- Alum addition to wastewater for the removal of phosphorus also produced a waste activated sludge which is easier to dewater. Also, when this sludge is combined with raw primary sludge, the resulting mixture again shows an increased ability to be dewatered (51).

- Vacuum filtration of iron sludge at the North Toronto plant showed the filter yield to increase from 11.08 to 23.14 kg/m^2/hr (2.27 to 4.74 lb/ft^2/hr). However, these data are not considered typical in view of the fact that an extremely low capture of solids was obtained. No data are available on sludge conditioning chemical requirements at this plant (72).

Conclusions

Some plants have experienced adverse effects on vacuum filter operation as the result of the addition of phosphorus-laden chemical sludges, while other plants have not. Unfortunately, there are no clear patterns apparent on which to base recommendations for "best" treatment. Treatment plants which are implementing phosphorus removal should conduct bench-top (or preferably pilot) tests, to determine the combination of phosphorus removal chemicals and sludge conditioning chemicals to obtain optimum vacuum filter performance. Experiments with various filter media are also suggested.

Some generalizations may be made as follows:

- Many plants reported operating their vacuum filters additional shifts to handle the additional chemical sludges generated.
- It is advantageous to thicken and condition the sludge prior to vacuum filtration in order to increase filter yield and filter cake solids concentration.
- Ferric chloride and lime are reported successful in conditioning alum sludges. Even with conditioning, however, it is common for alum sludges to have a lower percentage of filter cake solids; alum sludges are more difficult to dewater.
- Polymers have been reported successful in conditioning iron sludges.
- Filtration of combined primary-secondary sludges is preferable to filtration of secondary sludges alone.
- Iron sludges are more corrosive, so system component materials should be selected for corrosion resistance.
- Lime addition for phosphorus removal greatly increases the amount of sludge generated, and system components should be sized to accommodate the anticipated increase. Lime scaling may be a problem on wetted surfaces, so maintenance access for cleaning should be provided.
- Many dramatic improvements in sludge filterability have been reported as resulting from the proper use of polymers during sludge treatment steps prior to filtration.

Reported changes in filter yield and filter cake solids showed a confused pattern, with some plants experiencing increased yields and solids, and others showing poorer results than were experienced with non-chemical sludges. Alum sludge particularly was singled out as often difficult to dewater.

DRYING LAGOONS

Introduction

Drying lagoons form a simple, low-cost sludge dewatering system which is limited in application to areas where large quantities of cheap land are available. Of plants which responded to the questionnaire survey and/or were field investigated during this study, drying lagoons were used by 15 out of 174 (9 percent). Lagoons were the third most common dewatering method for chemical sludges, following drying beds and vacuum filtration.

In evaluating the impact of phosphorus removal by chemical addition on drying bed operation, we must consider how sludge drying rates will be affected and how plants may respond to the need for additional drying bed capacity for increased sludge volume. Sludge drying rate is important because sludges that take longer to dry have lower solids loading rates, and they require more space/ton of dry solids. Sludges that dewater slowly also tend to be the cause of odor problems. The need for additional drying lagoon capacity is often problematical for plants because of the expenses involved in making more space available.

Questionnaire Survey and Case Studies

As shown in Table 24, relatively little information on lagoon operation was obtained from the surveyed plants. Evidently, little monitoring of sludge drying rates, dried sludge TS concentrations, etc., is performed. The comments in Table 24 indicate that often lagoons were used as stand-by or supplementary operations in addition to other dewatering methods, and since they were not relied on closely, they were not monitored. Furthermore, at some of the plants, lagoons were treated as storage areas rather than as a dewatering method. In some cases, the plants did not expect to completely fill their lagoons for several years, so they had not been particularly concerned over control of dewatering rates or over sludge removal.

On the other hand, Table 24 also contains comments from plants which had run out of lagoon space and had been forced either to clean out the filled lagoons to make space available, or to purchase land and construct new ones. Both of these alternatives were considered expensive by the plants.

Some of the plants mentioned that their need for additional lagoon space was created by the additional sludge generated by phosphorus removal. Other adverse impacts of phosphorus removal which were mentioned included slower sludge drying rates, failure of the sludge to dewater, and odor problems. In contrast, there were other reports of chemical sludges which settled readily in the lagoons, formed clear supernatants, and posed no odor problems.

Conclusions

Drying lagoon operation is often made more expensive and difficult by the need to handle chemical sludges because of greater sludge volumes and slower drying rates. Plants must be prepared to provide more space for the extra sludge volume by constructing new lagoons or by cleaning out existing ones which

TABLE 24. IMPACTS OF CHEMICAL SLUDGES UPON DRYING LAGOON PERFORMANCE AS REPORTED IN QUESTIONNAIRE RESPONSE

Plant Location	Type of Sludge Treated	Impacts of Chemical Sludge	Performance with Chemical Sludge
Virginia, Minnesota	Lime-primary and secondary (gravity thickened) from an activated sludge plant	When phosphorus removal was begun with lime, the plant's digester and drying beds were abandoned. It was believed that the lime sludge could not be fed to the digester without causing problems. A gravity thickener and drying lagoons were chosen instead. The sludge was limed at the thickener to reduce odors, using 200 lb/day lime at \$0.15/lb.	An 8 to 10 percent TS sludge was generated by the thickener. But the sludge would not dry in the lagoons. The sludge also emitted a very offensive odor after a period of time even when heavily limed. The lagoons were abandoned and the sludge is hauled at 8 to 10 percent TS to a disposal site.
Ottawa, Ontario (Green Creek)	Alum-primary (anaerobically digested) from a plant with no secondary treatment	Alum addition created about a 90 percent increase in sludge volume, or approx. an additional 378 m^3/day (100,000 gpd). This necessitated the construction of 28.5 acres extra lagoon space (for a total of 56 acres) at a cost of \$690,000.	
Berrien Springs, Michigan	Iron-tertiary from an activated sludge plant	The chemical sludge is pumped directly to a holding lagoon which has a 20-yr design capacity. Organic sludge is treated separately.	There have been no odor problems.

(continued)

TABLE 24 (continued)

Plant Location	Type of Sludge Treated	Impacts of Chemical Sludge	Performance with Chemical Sludge
Coldwater, Michigan	Iron-primary and secondary (anaerobically digested) from a trickling filter plant	The sludge drying lagoon was cleaned out shortly after chemical addition for phosphorus removal was initiated to provide additional sludge drying space. This was necessitated by a slower sludge drying rate as well as greater quantities of digested sludge.	Sludge is sent to the lagoons on a sporadic basis at times when the drying beds are full and there is a need to make room in the digesters. No formal records have been kept as to the characteristics of the sludge in the lagoon.
North Madison, Ohio	Lime-tertiary from an activated sludge plant	The plant has had phosphorus removal since it started operation in November of 1974. Therefore, it is not possible to compare operation with and without chemical sludge.	The lime sludge is pumped to 2 lagoons which are 61 m by 30.5 m (200 ft by 100 ft). The sludge settles and a clear supernatant is decanted by a pipe at the 1.5-m (5-ft) water height. Cleaning the lagoons is an expensive problem. Both are full. $7,000 was paid in 1976 to a private contractor to clean 1/4 of a lagoon.
Sturgis, Michigan	Alum-tertiary from a trickling filter plant		The chemical sludge is not combined with primary and secondary sludges. It is pumped separately to so-called "dry lagoons." When a lagoon is filled it is allowed to dry and then scarified. This has not been done yet as a lagoon has not been filled.

(continued)

TABLE 24 (continued)

Plant Location	Type of SLudge Treated	Impacts of Chemical Sludge	Performance with Chemical Sludge
Ludington, Michigan	Lime-tertiary from an activated sludge plant		The chemical sludge is not combined with primary and secondary sludge. It is pumped to lagoons which are not expected to be filled for 20 yrs.
Watt's Creek STP, Shirley's Bay, Ontario	Alum-secondary and primary (anaerobically digested) from an activated sludge plant	With alum addition the raw sludge volume has increased 65 percent or more. The digested sludge solids concentration decreased from 5.5 percent to 3 percent, meaning an even greater increase in digested sludge volume. Increased holding lagoon space is planned to accommodate the extra sludge. An additional sludge hauling cost of $25,000 is anticipated for 1976.	
Niagara Falls, Ontario	Iron-primary (anaerobically digested) from a plant with no secondary treatment	The sludge from the digesters has a lower TS concentration and the volume is greater. The lagoons must be filled to deeper depths to accommodate the extra volume. The sludge takes longer to dry in the lagoons and does not reach the same solids concentration as before.	

have been used simply as storage areas. They must also anticipate the possibility of slower sludge drying rates, causing an even further demand for greater capacity. Where capacity is unavailable, alternative means of sludge handling such as liquid sludge hauling or mechanical dewatering will have to be considered.

PRESSURE FILTRATION

Introduction

Pressure filters have long been used in Europe to process difficult-to-dewater sludges. During the last five years, there has been a substantial increase in the use of pressure filters in the U.S. Improvements in equipment and greater quantities of difficult-to-dewater sludges, such as iron and alum sludges, account for the increase.

For conventional (non-chemical) sludges, the process produces a drier filter cake than either vacuum filtration or centrifugation. Total solids concentrations of 40 to 60 percent are generally achieved with undigested primary and secondary sludges, and concentrations of 40 to 45 percent TS are typical for digested primary and secondary sludges. Sludge thickening is usually required prior to filtration for all but primary sludges. Proper sludge conditioning is necessary in order to achieve the best filter results.

Evidently, the only conditioning method utilized at present is chemical conditioning. Polymer, lime, ferric chloride, fly ash, lime kiln flue dust, and sludge incinerator ash are the conditioners used. Frequently, a combination of approximately 5 percent (of the dry weight of the sludge) $FeCl_3$ and 10 percent lime is employed. The use of an additional 100 to 250 percent ash can usually increase filter cake dryness by about 5 percent.

For conventional sludges, the cost of dewatering by pressure filtration is higher than the cost of vacuum filtration or centrifugation. However, the drier cake produced may result in cost savings in the downstream disposal processes which offset or exceed the higher cost. The highest operation and maintenance expenses for pressure filtration are typically the cost of labor and the cost of chemical conditioners.

When dewatering chemical sludges, both the performance characteristics of the process and the energy, labor, and chemical requirements are altered. Consequently, the costs associated with pressure filtration, and also with the downstream disposal processes, are changed. It is desirable to determine whether chemical sludges enhance or decrease the cost-effectiveness of pressure filtration in relation to other dewatering methods.

Questionnaire Survey

A total of eight (46 percent) of the plants responding to the questionnaire survey reported that they have pressure filters. Only six questionnaire respondents commented upon the performance of pressure filtration at their plant or the impacts of the chemical sludge. Table 25 indicates the sizes of the six plants and the sludge treatment/disposal methods which were used. Thickening and digestion of the raw sludge were common. Two plants practiced incineration of the dewatered filter cake. A variety of sludge conditioners were used prior to dewatering. A single plant used an organic polymer. The others, including the two plants with incineration, used inorganic conditioners.

Table 26 presents data and comments from the questionnaires pertaining to pressure filter performance with conventional and chemical sludges. Four of the plants used plate and frame-type pressure filters and two used unusual types of filter processes. The plants, which had iron and lime sludges, reported that, when handling the chemical sludges, they needed smaller dosages of chemicals to condition the sludges before dewatering. These plants had filter cake solids concentrations of 40 to 50 percent TS.

Case Studies

The case study of Brookfield, Wisconsin, in Appendix H, studies the impacts of an iron sludge on the plant's plate and frame pressure filter. Phosphorus removal with ferrous sulfate resulted in an increase in the mass of sludge generated which was offset by a decrease in the weight of chemicals needed for conditioning. Although roughly 30 percent more sludge TS was generated during phosphorus removal, the dry weight of the filter feed, consisting of both sludge and conditioning chemicals, was greater by only 15 percent. The dosage of incinerator ash for conditioning was lowered from 85 percent to 60 percent. The $FeCl_3$ dosage also decreased somewhat, to a dosage of 6 to 8 percent . The lime dosage remained constant at 17 percent.

The chemical sludge did not alter the TS concentration of filter cake, which remained at 43 percent. The VS fraction of cake TS decreased only slightly, from 33.6 to 32.6 percent. Two positive impacts of the chemical sludge were a decrease in the average length of a filter run from 2.83 to 1.73 hr/run and an increase from 75 to 90 percent in the filter cake solids recovery.

The costs of sludge treatment and disposal at Brookfield were lower for the iron sludge. A decrease of about \$1.47/t (\$1.33/ton) occurred because of decreases in the amounts of chemical conditioners and electricity used by the pressure filter.

TABLE 25. SLUDGE TREATMENT/DISPOSAL METHODS USED BY PLANTS PRACTICING PRESSURE FILTRATION

Plant Location	Size (m^3/day (mgd))	Thickening	Digestion	Chemical Conditioning	Incineration	Disposal
Saline, MI	4,807 (1.27)	None	Anaerobic digestion	$FeCl_3$, Lime, Fly Ash conditioning	None	Sanitary landfill
Westfield, NY	5,678 (1.5)	Gravity thickening	Aerobic digestion	Polymer conditioning	None	Sanitary landfill
Kenosha, WI	75,700 (20)	Flotation thickening of WAS	Anaerobic digestion	$FeCl_3$, Lime conditioning	None	Cropland
Coloma, MI	4,542 (1.2)	Gravity thickening	None	Lime Kiln Flue Dust conditioning	None*	Cropland
Brookfield, WI	9,084 (2.4)	Gravity thickening	Aerobic digestion of WAS	$FeCl_3$, Lime, Incinerator Ash conditioning	Multiple hearth incineration	Sanitary landfill
Hatfield Township Colmar, PA	8,706 (2.3)	Gravity thickening	None	Lime conditioning†	Multiple hearth incineration	Sanitary landfill

* When the plant began operation in 1973, multiple hearth incineration was used and chemical conditioning was performed with lime and incinerator ash.

† In the past, $FeCl_3$ was also used for chemical conditioning.

TABLE 26. IMPACTS OF CHEMICAL SLUDGES UPON PRESSURE FILTER PERFORMANCE AS REPORTED IN QUESTIONNAIRE RESPONSE

Plant Location	Type of Sludge Treated	Impacts of Chemical Sludge	Performance with Chemical Sludge
Saline, Michigan	Iron-primary and secondary from a trickling filter plant	The entire plant was expanded and the pressure filter was installed when phosphorus removal was begun. Therefore there is no comparison between pressure filter operation with and without the iron sludge	The digested sludge at 9 percent TS produces a filter cake which averages 41 percent TS. The cake vs fraction averages 62 percent of TS.
Westfield, New York	Aluminum-secondary from a two-stage long-term activated sludge plant with no primary treatment	The plant incorporated phosphorus removal since it went into operation. Therefore, there is no data on the impacts of the aluminum sludge.	The plant is using an unusual type of belt filter press which is similar to a vacuum filter with a belt squeezing the sludge to the cloth. The sludge is difficult to dewater because it is all waste activated. The filter cake TS concentration averages 11.5 percent. Polymer conditioning of the aluminum sludge was found to be more successful than conditioning with $FeCl_3$ and lime.
Kenosha, Wisconsin	Iron-secondary and primary from a conventional waste activated sludge plant	The amount of $FeCl_3$ required for chemical sludge conditioning was reduced from 5 percent to 3½ percent. (The lime dosage is 18 to 22 percent.)	The digested sludge at 4.8 percent TS produces a filter cake which averages 40 percent TS.

(continued)

TABLE 26 (continued)

Plant Location	Type of Sludge Treated	Impacts of Chemical SLudge	Performance with Chemical Sludge
Coloma, Michigan	Lime-primary and alum-secondary from a trickling filter plant	The use of lime increases the TS concentration of the thickened sludge. It also decreases the requirement for lime kiln flue dust for chemical conditioning.	The estimated average filter cake TS concentration is 40 percent. The plant used incineration when it began operation but it was very expensive because of the low VS content of the sludge. Due to the large amounts of lime, lime flue dust and incineration ash used for phosphorus removal and sludge conditioning the VS fraction of filter cake TS was only 20 to 30 percent.
Brookfield Wisconsin	Iron-secondary and primary from a contact stabilization activated sludge plant	See the following case study discussion of Brookfield.	
Hatfield Twp, Colmar, Pennsylvania	Lime-primary secondary and alum-tertiary from a complete-mix activated sludge plant	The amount of sludge generated when a plant expansion including phosphorus removal took place was much greater than expected. At that time the plant was practicing vacuum filtration and incinerating a 22 percent TS filter cake. The incinerator loading rate was too low to keep up with the amount of sludge generated. In response to this problem, the vacuum filter was	The plant has an unusual type of filter press which has been adapted from the old belt vacuum filter. Sludge is actually vacuum filtered and then carried by the long vacuum filter belt to a diaphragm-type filter press. The system produces a 45 to 50 percent TS filter cake which has a VS fraction ranging from 27 to 40 percent of TS. The filter contains 299 mg/ℓ SS, 16 mg/ℓ VSS, 917 mg/ℓ COD, and

(continued)

TABLE 26 (continued)

Plant Location	Type of Sludge Treated	Impacts of Chemical Sludge	Performance with Chemical Sludge
		adapted to a diaphragm-type filter press. They produced a 45 to 50 percent filter cake. This allowed a higher incinerator loading rate. The addition of lime for phosphorus removal is believed to be reducing the lime requirement for chemical conditioning.	872 mg/ℓ BOD.

Literature

An evaluation of lime sludge dewatering by centrifugation and pressure filtration was made at the Holland, Michigan, wastewater treatment plant (43). The design of the plant included addition of lime to the primary step for phosphorus removal and aluminum addition to the activated sludge treatment step for effluent polishing. Sludge treatment was provided by primary sludge degritting, primary and waste activated sludge gravity thickening, centrifugation and fluidized bed incineration.

With this system, a 30 to 32 percent TS centrifuge cake was incinerated, and the cost of fuel (natural gas) for incinerator was found to be a major operational expense at $26.92/t of dry cake solids (24.42/ton) in 1974. Pressure filtration was then considered as a means of reducing the fuel requirement by increasing the sludge cake TS concentration. The results of a pilot test are shown in Table 27. A 0.093 m^2 (1 ft^2) pilot press with three plates and two chambers was operated at 125 psig pressure for this test.

TABLE 27. PILOT FILTER PRESS TEST AT HOLLAND, MICHIGAN (43)

Total Solids in Feed % TS	Filtration Rate (gal/hr/ft^2) (l/hr/m^2)	Cake Conc. %	Pressure kg/m^2 (psig)	Polymer kg/t (lb/ton)
11.2	178 (4.33)	41.54 with thickness 1.25"	8.8 (125)	1.5 (3 lb/ton) Hercules 859-C

Based on the pilot test data, estimates of the performance and cost of full-scale pressure filtration were made. It was assumed that polymer conditioning of the sludge would be used. It was estimated that a pressure filter operating at 225 psig would produce a cake of 45 percent TS. The filtration cycle would be 1 hr based on a feed sludge TS concentration of 8 percent. With the incinerator operating at a freeboard temperature of 827°C (1,520°F), and with a 28 percent volatile filter cake, the fuel cost would be $12.03/t ($11.15/ton).

Pressure filtration was compared to dewatering with a centrifuge or with the existing vacuum filters at the Hatfield Township, Pennsylvania, plant (31). This plant was described in Tables 25 and 26. A mixture of lime-primary, waste activated and alum-tertiary sludges was dewatered by the three different methods under similar operating conditions. The pressure filter produced a drier cake with greater recovery of solids than either the vacuum filter or the solid bowl centrifuge. A 41 percent TS filter cake was achieved with sludge conditioning with 15 percent lime.

Conclusions

There is insufficient data to make conclusions about the dewatering of alum and waste activated sludges with pressure filters. But there are indications that the results are not as good for these sludges as they are for iron or lime sludges and for primary or combined primary and secondary sludges.

The results of the questionnaire survey, case studies, and literature search have not included any mention of adverse impacts of iron or lime sludges on pressure filtration. These sludges dewater well on pressure filters, with filter cake TS concentrations of 40 to 50 percent and often with reduced chemical requirements for sludge conditioning. Because of the reduction in chemical costs, the cost-effectiveness of this dewatering method alternative can be favorably affected.

CENTRIFUGATION

Introduction

Sewage sludge dewatering by centrifuges began to evolve in the 1950's. Their use has been gradually increasing due to better engineering and materials of construction. Of plants which responded to the questionnaire survey and/or were field investigated during this study, centrifuges are used by 6 out of 174 (3 percent). Several additional plants have conducted promising pilot tests with centrifuges and are considering their future full-scale use to help solve sludge management problems.

Proper centrifuge design is complex, and it is assumed that the reader has a basic knowledge of centrifuge design and operational factors. Essentially, centrifuges have the same objectives as vacuum filters, e.g., maximum solids capture, maximum yield per unit of energy consumed, high solids concentration in the sludge cake, and low total cost. In addition, since centrifuges are high maintenance units, it is desirable to have good resistance to abrasion and long-living moving parts.

Most centrifuges dewatering municipal sewage sludge are of the continuous, solid bowl, conveyor type. A relatively new, low-speed, solid bowl centrifuge which was developed in Europe is now available in the United States and is reported to have some significant advantages. Basket centrifuges are also available.

Because each sludge and treatment plant situation is unique, the information presented in the following subsection is not intended to provide universal answers to specific plant problems. Rather, the information is intended to be helpful to those seeking possible avenues of further investigation. The best method of predicting the true performance of full-scale centrifuges is experiments with pilot units on the actual sludges to be treated. These pilot units are available from manufacturers who have developed scale-up design factors for their units.

Literature

There are hundreds of literature references pertinent to centrifuging of sewage sludge. This study concentrated upon those which discussed centrifuging of phosphorus-laden chemical sludges. Generally, the literature favorably evaluates centrifugation of chemical sludges (particularly lime sludges), as evidenced by the selected summaries below.

In connection with the lime treatment nutrient removal program at the Newmarket, Ontario WPCP plant, a 2-mo raw sludge dewatering study using a Sharples solid bowl centrifuge was undertaken (65). Centrifuge variables, such as sludge feed concentration, sludge feed rate, sludge pH, and speed differential were examined. The project was successful and the optimum results produced from centrifuging raw lime primary sludge were as follows:

(a) 98 to 99 percent SS capture

(b) Polymer dosages of less than 0.5 kg/t (1 lb/ton) of dry solids

(c) Consistent centrate qualities of less than 700 ppm SS

(d) Sludge cake of 27 to 34 percent TS

(e) Feed to centrate, total phosphorus reduction of 95 percent

(f) Centrate total phosphorus levels in the order of 10 to 30 ppm.

Using the Sharples P3000 Super-D-Canter centrifuge, solids recovery and cake dryness could be attained to some degree with unconditioned lime sludge. Augmentation with polyelectrolytes showed increased overall solids recovery and vast improvements in centrate quality.

The Orange County Water District practiced chemical clarification of secondary effluent with lime at a pilot wastewater reclamation facility. They removed lime sludge from the pilot tertiary clarifier and treated it in a small pilot thickener and a 15-cm (6-in) solid bowl centrifuge. Sludge was removed from the clarifier at 1 to 2 percent TS and could be thickened in the pilot thickener to 10 to 20 percent TS. The centrifuge data showed that the thickened sludge could be dewatered to over 50 percent TS in a Sharples P-600 Super-D-Canter centrifuge.

Sludge dewatering studies (2) at the Portage Lake Water and Sewer Authority, Michigan, concluded that alum-biological sludges were amenable to conventional sludge treatment processes. Of the various concentrating and dewatering schemes investigated for the alum-biological sludge, two-stage aerobic digestion followed by basket centrifugation appeared to be the most economical. However, because of the short distance to the disposal site, direct trucking was estimated to produce a 10 percent savings over centrifugation.

In a detailed report by Martin and Nardozzi (43), performance and cost information for lime-biological sludge treatment at Holland, Michigan, was provided. The city of Holland used lime precipitation of phosphorus in the primary clarifier, with $AlCl_3$ addition to the flocculator prior to biological treatment. The thickened, combined primary solids and waste activated sludge was fed to solid bowl centrifuges prior to incineration. The centrifuges produced a cake of 30 to 32 percent total dry solids, with 97 percent solids capture.

The polymer (Hercules 869-C) dosage rate was 2.0 to 2.5 kg/t (4 to 5 lb/ton) of dry solids. When polymer was used for conditioning, more of the finer, less dense particles were captured. The data revealed that the highest cake solids were achieved when no polymer was used. As the capture increased with polymer addition, the cake became wetter, because the fines and associated water were captured. Even so, polymer addition was advisable because a cleaner centrate ensured a lighter recirculating load and a minimum sludge age in the plant.

The EPA, in 1975, conducted pilot scale investigations (26) of dewatering chemical sludges with vacuum filters and centrifuges. The investigations were made at various midwestern plants and were plagued by mechanical difficulties with the trailer-housed pilot equipment. The most significant centrifuge results were the characteristics of the centrate from various pilot-scale centrifugation runs. These data, shown in Table 28 provide a tool with which to study the impact the various sidestreams could have on plant operation and performance. Note particularly the heat-treated iron sludge from the Midland, Michigan, plant which demonstrated a high loading rate and cake solids concentration, but also showed a high solids and organic load in the centrate. This centrate returned to the plant could cause operational difficulties.

The EPA study further concluded that:

Increasing the polymer dosage decreased the cake solids.

TABLE 28. CENTRATE CHARACTERISTICS FROM VARIOUS SCROLL CENTRIFUGATION RUNS (26)

Sludge	Bowl Speed, rpm	Loading Rate Kg/hr (lb/hr)	Cake Solids %	Centrate Solids %	Polymer	Solids Recovery %
Beaver Creek	5000	0.671 (1.48)	4.3	0.01	yes	100
Waste Activated		1.26 (2.77)	5.1	0.02	yes	95.8
Alum-Sludge		1.61 (3.55)	5.42	0.03	yes	95.3
Beaver Creek	5000	1.64 (3.62)	5.79	0.01	yes*	94.2
Aerobic Digested		2.42 (5.32)	10.9	0.04	yes	99.7
Alum-Sludge		3.20 (7.05)	6.32	0.16	yes*	90.0
		4.86 (10.7)	6.37	0.33	yes*	84.6
		7.44 (16.4)	15.2	0.09	yes	99.2
Midland	4000	19.2 (42.3)	33.3	4.5	no	82.9
Heat Treated	4500	24.7 (54.4)	49.0	9.6	no	97.1
Iron-Sludge						

* Added to sludge ahead of centrifuge.

This was also borne out in separate Canadian studies.

- Much better results are achieved from polymer addition at the centrifuge, rather than ahead of the centrifuge. Higher solids recovery, better centrate quality, and higher cake solids concentration were achieved by polymer addition at the centrifuge.
- Increasing the feed rate resulted in decreasing the solids recovery using aerobically digested alum-secondary sludge.
- For similar conditioning, both solids recovery and cake solids concentration are generally less for chemical sludges than for conventional sludges.

Questionnaire Survey and Case Studies

As shown in Table 29, there was relatively little information obtained from the questionnaires pertinent to centrifugation of chemical sludges. More complete information was obtained from the Port Huron, Michigan, and South Bend, Indiana, case studies in the appendices.

The Port Huron case study report examines some of the factors influencing centrifuge dewatering of an alum sludge:

- Centrifuge loading rate was higher with thermal sludge conditioning than with polymer conditioning.
- The dosage of polymer required for conditioning was higher when the feed sludge TS concentration was low as the result of high activated sludge wasting rates.
- The TS concentration of the dewatered cake was higher with thermal conditioning than with polymers. The cake TS concentration apparently did not vary as the result of changes in the feed sludge TS concentration if polymer dosages and feed rate were adjusted.
- The average centrate TS concentration was lower during thermal conditioning while the volatile fraction of TS in the centrate was higher.
- The TS capture could be maintained at at least 80 percent by the operator. It was necessary for the operator to check the centrate of each centrifuge separately, several times each day. He then made separate adjustments of feed rate and polymer dosage for each centrifuge.

It was not possible to judge the impacts of alum addition for phosphorus removal on centrifuge performance at this plant because alum had been in use since the plant was upgraded. However, it seemed likely that the quantity of waste activated

TABLE 29. IMPACTS OF CHEMICAL SLUDGES ON CENTRIFUGE PERFORMANCE AS REPORTED IN QUESTIONNAIRE RESPONSE

Plant Location	Type of Sludge Treated	Summary of Comments
Lower Allen Township, Camp Hill, Pennsylvania	Lime-primary and secondary (gravity-thickened) from an activated sludge plant.	There was a great increase in sludge mass. The centrifuge dewatering operational costs greatly increased. The lowered volatile content in the sludge cake increased the subsequent incineration operational costs. Had to reroute centrate return flow so that instead of being added to the thickeners, it was sent to the head of the plant because the centrate return was causing odor and floating solids problems in the thickeners. They had difficulty finding polymers that were effective in conditioning the sludge by adding to the centrifuges. The dewatered cake contains 30 percent TS.
Hilton, New York	Alum-secondary and primary (gravity-thickened) from an activated sludge plant.	Centrifuge operation was not significantly affected by addition of the chemical sludge. Polymer continued to be added at a rate of 6 to 7-1/2 kg/t (12 to 15 lbs/ton), and cake solids remained at about 19 percent.

relative to primary sludge was greater because of alum addition to the aeration basins. Greater quantities of waste activated sludge meant higher polymer conditioning requirements at the plant.

The South Bend case study report describes centrifugation of a lime/iron sludge resulting from tertiary addition of both lime and iron to secondary effluent. The sludge was gravity thickened and fed to the centrifuges at 10.2 percent TS. The cake produced contained 43 percent TS. There were problems with maintenance of the centrifuges. The heavy lime sludge was responsible for breaking and frequent equipment overhauls.

Conclusions

The information which was presented indicates that thickened lime-primary, or primary plus secondary, sludges can be dewatered to 27 to 34 percent TS in a centrifuge. Centrifuge dewatering of thickened lime-tertiary sludges apparently produces an even drier cake of 40 to 50 percent TS. At one plant, a combined alum-secondary and primary sludge was gravity thickened and centrifuged to 19 percent TS. Centrifuge performance depends heavily on choosing the right conditioning method. Polymers are usually used. The point of polymer addition is important, some experiences indicating that the polymer should be added at the centrifuge rather than before it. The polymer dosage is also important and should be adjusted as feed sludge consistency and TS concentration vary.

Iron and aluminum chemical sludges often require higher conditioning polymer dosages and lower feed rates than conventional sludges. Poor centrate quality and low cake TS concentrations result otherwise. Closer operator attention may also be necessary with iron and aluminum sludges because of inconsistent gravity thickener performance, meaning variations in feed sludge quality.

SECTION 9

REDUCTION OF CHEMICAL SLUDGES - INCINERATION

INTRODUCTION

Incineration uses thermal energy to provide a major reduction in the weight and volume of solids requiring disposal. The major factor in evaluating incineration performance and costs is the composition of the sludge feed. Sludge composition, i.e., moisture, volatile solids, and inert contents, influences feed rate capacity and auxiliary fuel requirements, among other variables, and, consequently, sludge incineration cost effectiveness. Due to their inert content, chemical sludges have been recognized as poor incineration feed (43 and 45).

QUESTIONNAIRE SURVEY

As previously shown in Table 10, 22 of the plants (13 percent) responding to the questionnaire survey reported that they incinerated their chemical sludges. Six plants surveyed indicated that chemical addition for phosphorus removal was having a significant impact on their incineration systems. Table 30 summarizes the experiences of these plants. All plants incinerated combined primary and secondary sludges. Lime, ferric chloride, and alum were used at two of the six respondents.

Problems with incineration encountered by the plants surveyed generally agreed with literature reporting the difficulties with incineration of chemical sludges. The lower TS concentration and lower volatile content of the phosphorus-laden sludges made auto-combustion of the sludge harder to attain resulting in soaring auxiliary fuel requirements.

Increased fuel requirements caused a few plants to abandon their incinerators and adopt other sludge handling methods, i.e., dewatering and hauling. In addition, the increased inert content of the chemical sludges put a burden on incinerator ash handling equipment and, in certain cases, clinker buildup within the incinerators resulted. Problems with clinkers resulted in periodic downtime for incinerator maintenance. The Wyandotte, Michigan, plant used primary additions of ferric chloride for phosphorus removal and reported clinker buildup within its multiple hearth incinerator. The clinkers plugged the multiple hearth dropholes and were identified as the source of increased rabble arm wear.

TABLE 30. IMPACTS OF PHOSPHORUS-LADEN CHEMICAL SLUDGES ON INCINERATOR PERFORMANCE AS REPORTED IN QUESTIONNAIRE RESPONSE

Location	Size-m^3/day (mgd)	Type of Sludge Treated	Impacts of Chemical Sludge
Warren, Michigan	123,000 (32.5)	Alum - secondary and primary from a conventional activated sludge plant	Higher gas costs for incineration resulted.
Wyandotte, Michigan	291,400 (77)	Iron - primary and secondary from a pure oxygen activated sludge plant	In the multiple hearth incinerator there has been a major problem with clinker buildup, causing more time to be spent cleaning up the drop holes in the hearth and more wear on the rabble arms.
Coloma, Michigan	4,500 (1.2)	Lime - primary and alum secondary from a trickling filter plant	The addition of chemicals for phosphorus removal has lowered the sludge volatile content to 20 to 30 percent, causing the need for continuous use of auxiliary fuel (natural gas).
Camp Hill, Pennsylvania	7,600	Lime - primary and secondary from a complete mix activated sludge plant	The multiple hearth incinerator was designed to recalcinate and reclaim lime; this step was eliminated due to economic considerations. With the lower volatile content of the sludge, the incinerator requires between 80 and 100 gal of auxiliary No. 2 fuel oil per dry ton of sludge.

(continued)

TABLE 30 (continued)

Location	Size-m^3/day (mgd)	Type of Sludge Treated	Impacts of Chemical Sludge
Sheboygan, Wisconsin	54,900 (14.5)	Iron - secondary and primary from a trickling filter plant	The incinerator requires an additional 12 to 15 gal/hr of fuel oil due to the lower volatile content of the solids.
Richardson, Texas	6,700 (1.78)	Alum - secondary and primary from a trickling filter plant	The plant originally used anaerobic digesters and drying beds for sludge disposal. With alum addition for phosphorus removal, the volume of sludge to be disposed of increased by 50 percent. Dewatering equipment was installed and an electric incinerator designed to incinerate the dewatered cake. Leaf tests were run on the sludge, and these showed the sludge could be dewatered to 20 to 25 percent dry TS with a 60 percent volatile fraction, and the dewatering equipment and electric incinerator were designed on this basis. However, in practice, the best results from dewatering were 15 percent dry TS and a 45 to 50 percent volatile fraction. This sludge was not judged to be suitable for incineration, and conditioning with lime, ferric chloride, and different polymers did not improve the system's performance.

CASE STUDIES

The case studies in the Appendices contain detailed information on the effects of chemical addition on incineration. The particular case studies dealing with this subject are Sheboygan, Wisconsin; Port Huron, Michigan; and Brookfield, Wisconsin.

The Sheboygan, Wisconsin, wastewater treatment plant operates a Dorr-Oliver FluoSolids incinerator. This fluidized bed incinerator was designed to handle a sludge feed of 25 percent TS with a 73 to 74 percent volatile fraction. With the addition of ferric chloride for phosphorus removal, the sludge feed TS concentration was 22 percent with a volatile fraction of 65 percent. These deviations from the design capacity of the incinerator unit resulted in a decrease in the incinerator's loading capacity and an increase in auxiliary fuel consumption. The decrease in capacity was judged to be approximately 121 kg (266 lb of dry TS/hr, and No. 2 fuel oil consumption was judged to increase by approximately 143 ℓ/t (34 gal/ton).

There have also been increased incinerator maintenance and repair problems at the plant due to slag formation and corrosion. On one occasion, four tuyeres blew out because several of them were plugged with slag. On another occasion, a pressure buildup in the reactor signaled a problem, and a large piece of slag was found blocking the exhaust line. After this experience, the operator performed a visual inspection of the exhaust line from the roof duct inspection port every 3 to 4 mo.

In addition, the plant manager suggested that the ferric chloride used for phosphorus removal was responsible for a high rate of metal corrosion in ducts, especially in the elbows of the scrubber system. Gradually, these parts were being replaced with stainless steel.

The Port Huron, Michigan, plant, which was handling an alum sludge, also operated a Dorr-Oliver FluoSolids incinerator. For a period of time, Port Huron thermally conditioned sludge with a Farrer System thermal conditioner prior to centrifugation and incineration. This step provided major savings on chemicals and sludge conditioning fuel oil. It also increased the solids loading to the incinerator. With thermal conditioning, the sludge burning rate reached 1,043 kg/hr. Dorr-Oliver has estimated that the incinerator capacity would be reduced to 590 kg/hr for non-thermally conditioned sludge.

The cost of sludge thermal conditioning amounted to $35.00/t ($31.74/ton) excluding equipment maintenance and repair supplies. Prior to thermal conditioning, chemical conditioning costs were $50.30/t ($45.66/ton). It therefore appeared that operation and maintenance of thermal conditioning at the Port Huron plant was cost effective as long as the cost of equipment maintenance and repair supplies was less than $15.30/t ($13.92/ton).

LITERATURE

A review of the literature on incineration of chemical sludges reveals a common set of problems: greater sludge volume, lower calorific value, increased moisture content, formation of clinkers due to iron or aluminum content, and increased inert content. The Canadian Environmental Protection Service has suggested solutions to handle these incineration problems (61).

Problem	Solution
Greater sludge volume	Increase capacity or run time
Lower calorific value	Increase auxiliary fuel
Increased moisture content	Improve dewatering or increase auxiliary fuel
Formation of clinkers	Decrease incineration temperature below fusion point
Increased inert content	Increase ash disposal capacity

In a wastewater treatment facility using large amounts of lime for phosphorus removal, consideration must be given to the possibility of recalcination of the chemical sludge. When lime is used, calcium carbonate and calcium hydroxyapatite are the major sludge components. The hydroxyapatite containing the removed phosphorus is fairly stable, but the calcium carbonate can be recalcined to recover lime. The decision to recalcinate lime should involve an economic study considering the cost of lime, the capital costs necessary to achieve recalcination, the additional operating costs of recalcination, and the cost of incinerator ash disposal. The latter item is one that very often is not considered in such an economic analysis, but reuse of 65 to 85 percent of the lime can significantly reduce the volume of ash generated from the incineration process. If costs associated with this final ash disposal are significant, it can have a marked effect upon the total economic picture.

The Holland, Michigan, wastewater treatment plant used primary addition of lime for precipitation of phosphorus and a small amount of Al^{+3} for effluent polishing (43). The primary and waste-activated sludges were combined and dewatered for incineration in a Dorr-Oliver FluoSolids reactor. In operating this fluidized bed reactor, Holland found that the sand in the bed served as a seed for chemical material to form agglomerates. The resultant bed material tended to be spherical, decreasing significantly the angle of repose. This resulted in backsifting of the sand through the distribution tuyeres into the windbox of the reactor. This problem was solved by plugging alternate tuyeres to increase pressure drops across the distribution plate and adding a gravel type base (olivine) to the top of the distribution plate. Backsifting was eliminated.

Another problem occurred at the feed tubes through which sludge entered the fluidized bed. The combined organic-chemical sludge baked into a solid mass when exposed to the heat of the reactor for a sustained period of time. To counteract this, the feed tubes to the reactor were shortened to decrease heat transfer from the bed. Normally, a plug was left in the reactor feed tube at the conclusion of daily operations. Because a plug would solidify under these conditions, it was necessary to replace the material with peat moss or digested sludge from a nearby digester. A tool was later developed to remove the baked plugs.

A study was undertaken at the Minneapolis and St. Paul, Minnesota, wastewater treatment plants to determine the relationship between chemical conditioner additions and incinerator auxiliary fuel requirements. Ferric chloride and lime addition at greater than 5 and 15 percent of the weight of the dry solids produced a drier feed cake, but it also increased the kg of water fed per kg of dry TS fed into the incinerator. Any increases in the kg water/kg dry TS fed into the incinerator also increased the auxiliary fuel requirements for cake incineration. The basic problem, then, in reducing auxiliary fuel requirements was seen to be one of reducing the ratio of water to sludge solids in the cake, and not one of reducing the moisture content of the cake on a percent TS basis.

In addition, the auxiliary fuel requirements were compared using alternative sludge conditioners. A polyelectrolyte-conditioned sludge was found to require significantly less auxiliary fuel than a lime-and-ferric-chloride-conditioned cake.

CONCLUSIONS

Various solutions to the inherent problems associated with incineration of phosphorus-laden chemical sludges have been presented in this discussion. Unfortunately, at some wastewater treatment plants modification of the incineration system to incorporate these solutions surpasses the cost effectiveness of the system. Therefore, a detailed cost analysis of the benefits of incineration modification versus other sludge handling methods must be made.

Generally, there are many indications that sludge handling systems designed to include incineration will not be cost effective for the treatment of chemical sludges unless they incorporate those sludge pretreatment technologies most capable of providing a filter cake of high volatile content and low moisture content. The advantages and disadvantages of these techniques were discussed at length in previous sections with reference to their effects on incineration.

SECTION 10

DISPOSAL OF CHEMICAL SLUDGES - TRANSPORT AND LAND DISPOSAL

INTRODUCTION

Table 10, which is based on the questionnaire responses of 174 plants practicing phosphorus removal by chemical addition, shows that virtually all of the plants practiced some type of land disposal of sludge. Even among plants practicing incineration, most periodically disposed of sludge cake on land or in landfills. Except in a few cases where sludge was lagooned or stockpiled at the plant, sludge transport to the final disposal was required. Transport costs sometimes comprised a very significant portion of the overall costs of land disposal. Sludge hauling by tank or dump truck was by far the most widespread method of transport. However, pipeline transport has successfully been used for both lime and iron sludges at South Bend, Indiana, as described in Case Study No. C (see Appendices).

QUESTIONNAIRE SURVEY AND CASE STUDIES

Because of the impacts of phosphorus removal on sludge production and treatment, plants were hauling more sludge in liquid and cake form to land disposal sites than before. The questionnaire responses verified that most of the plants which historically practiced sludge hauling had to haul greater volumes of sludge after phosphorus removal for numerous reasons:

1) Additional raw sludge solids were produced.
2) The raw sludge TS concentration was lower.
3) Sludge thickening was adversely affected.
4) Digester solids-liquid separation was poorer.
5) The dewatered cake contained more moisture.

At Manitowoc, Wisconsin, for example, alum addition increased sludge volume so that it was necessary to incorporate into daily operation the use of a 9.5-m^3 (2,500-gal) tank truck as well as the regular 11.3-m^3 (3,000-gal) truck. The capital cost of the truck was $43,382.00 in 1976, and the estimated operating cost was $17,590.00 in 1977.

At the Port Weller plant in St. Catherines, Ontario, a 26 percent increase in sludge volume with alum addition caused a 23 percent increase in the annual cost of liquid sludge hauling. Kingston, Ontario, reported that its sludge mass doubled with iron addition, and its liquid sludge hauling effort was 60 percent greater. Several other plants, ranging in size from 1,510 to 51,100 m^3/day (0.4 to 13.5 mgd), reported increases of $5,000.00 to $25,000.00 for hauling in 1976 due to iron or alum addition.

The questionnaire responses also indicated that many plants which historically have not hauled liquid sludge or sludge cake found it necessary to start doing so after phosphorus removal because:

1) the chemical sludge was unsuitable for the usual treatment processes, or

2) the quantity of chemical sludge exceeded the capacity of existing treatment facilities.

Liquid sludge hauling was frequently used by plants with inadequate dewatering and incineration capacity to dispose of extra sludge resulting from phosphorus removal. At Hatfield Township, Pennsylvania, for example, the pressure filter and incinerator could not handle all of the plant's sludge even when run for 24 hr/day, 7 days/wk, so liquid sludge was hauled periodically by a contractor. Other plants may have had the ability to mechanically dewater the extra sludge, but they preferred to haul it as liquid in order to avoid running the equipment longer and hiring more personnel. For instance, at Lakewood, Ohio, the plant could not handle the additional sludge volume generated by alum addition on its schedule of vacuum filter and flash dryer operation for one shift each day. Rather than extending filter and dryer operation to two shifts per day and hiring three more employees, the plant decided to have sludge hauled from the digesters at $132.30/t ($120.00/ton). Details on costs and the effects on plant performance are contained in Case Study No. F (see Appendices).

The questionnaire responses revealed only two plants -- Virginia, Minnesota, and Coloma, Michigan -- which were forced to haul all of their sludge to land disposal after phosphorus removal, abandoning their existing dewatering or incineration facilities. At the Virginia plant, anaerobic digestion and drying beds were used for sludge handling previous to phosphorus removal. In anticipation of phosphorus removal with lime, they were replaced with a gravity thickener and lagoons. The sludge refused to dry in the lagoons, so they were abandoned and liquid sludge was then hauled to a disposal site. This involved hauling sludge three times/wk as opposed to cleaning the drying beds twice each year. The plant was planning to purchase a tandem or triaxle tank truck for $40,000.00 to haul 189 m^3 (50,000 gal) of liquid sludge per week, and a heated enclosure for the truck and piping changes for $40,000.00. The estimated operating costs are $4,500.00/yr.

At Coloma, pressure filtration and incineration of sludge were practiced when the plant went into operation. Phosphorus removal with primary addition of lime and secondary addition of alum was begun a year later. Because of the large volumes of lime and flue dust used for sludge conditioning, the volatile content of the sludge was only from 20 to 30 percent, and incinerator auxiliary fuel consumption was high. Flower farmers in the area were contacted, and they agreed to take all of the sludge for its ability to improve the sandy, low pH soil on their farms.

Phosphorus removal can reduce the number of suitable land disposal alternatives which are available to a particular plant. For instance, sludge TS concentration of at least 20 percent is usually needed for landfilling of sludge along with other materials such as municipal solid waste, and at least a 40 percent TS concentration is necessary for landfilling of sludge alone. Iron and aluminum sludges are generally more difficult to dewater to these concentrations than conventional sludges with many of the dewatering methods available. Some lime sludges can easily be dewatered to 40 percent TS, but the plant must have the equipment to handle the large quantities of sludge which are generated with lime. Furthermore, there must be room in the landfill for the greater quantities of sludge which usually result from phosphorus removal. At Toledo, Ohio, for instance, it was necessary to switch from landfilling of vacuum filter cake to cropland application because of the additional volume of cake produced during phosphorus removal with ferric chloride.

LITERATURE

As with conventional sewage sludges, there is concern over the beneficial or detrimental effects of chemical sewage sludge application to croplands. There is special concern with regard to chemical sludges because concentrations of nutrients, metals, and organic and microbial constituents can be higher in these sludges than in conventional sludges. The use of lime, aluminum salts, or iron salts for phosphorus removal precipitates most of the metallic cations contained in the wastewater as well as the bulk of the phosphorus. These metals and the iron, aluminum, or calcium of the precipitating chemicals are concentrated in the sludge. A limited amount of investigation has been conducted into the effects on crop yield, heavy metal toxicity to crops and animals, and groundwater quality of applying these sludges to croplands. Of primary concern has been the state of heavy metals in sludge and the possibility of resolubilization in the soil. Resolubilization makes the metals available to be concentrated in crops and leachate, presenting the possibilities of toxicity to crops and aquatic life and accumulation in the human food chain and drinking water supply. Since resolubilization is most likely to occur in low pH soils, it is expected that lime sludges will present the least problem because of their tendency to raise the soil pH (72).

Kirkham and Dotson studied the growth of barley which was irrigated with liquid sludges produced with the addition of ferric chloride and alum to raw wastewater (39). The barley was grown in pots of loam soil for 4 mo. The chemical sludges did not limit the growth of the barley compared to plants fertilized with inorganic fertilizer or non-chemical sludge.

Chawla, et al., (15) conducted 1-yr lysimeter investigations using alum, iron, and lime sludges from various Ontario treatment plants. They planted orchard grass in loamy sand and silt loam amended with sludge. They found that grass yields comparable to fertilizer treatment could be achieved at the highest application rates (900 kg/ha nitrogen) of lime and iron sludges but that alum sludge produced much lower yields. Cumulative effects of higher zinc equivalent, higher aluminum contents and low potassium amounts in alum sludges might have retarded the plant growth. Very high concentrations of petroleum hydrocarbons in the alum sludge possibly caused some plant toxicity resulting in lower yields.

Increasing the rates of sludge application increased the nitrogen content of the plants; phosphorus did not show any meaningful trend; potassium contents decreased with increasing rates of sludge addition. Nickel, copper and zinc concentrations in plant tissues were reasonably low and similar to the control and fertilizer treatments. One year after sludging, no definite chemical changes in the soil systems were observed. Decline in pH values under highest rates of iron and alum sludges suggested a trend toward increased mineralization and acidification due to a lack of calcium.

Chawla and others (14) also conducted an intesive study of the biochemical characteristics of chemical sewage sludge from various municipal plants in the province of Ontario. The study pointed out that sludge from sewage treatment plants must be characterized on an individual basis to determine if possibly hazardous concentrations of heavy metals or other contaminants exist. The sludges studied varied widely in their biochemical compositions. It should be remembered that high levels of specific metals in sludge are virtually always attributable to specific industries in the area served by the plant.

Where heavy metals are concentrated to a considerable degree in sludge, Scott and Horlings (58) suggest that they can be easily removed by extraction with dilute acid. They experimented with anaerobically produced sludges, and found that both metals and phosphate can be extracted from thickener underflow sludge or dewatered filter cake. The products of this extraction can be further treated by other conventional means to recover phosphate or specific metallic components.

CONCLUSIONS

Hauling sludge liquid or cake to a land disposal site has been a common solution to many of the sludge processing problems created by phosphorus removal. This has been the case even though phosphorus removal generally increases sludge hauling costs because of greater sludge volumes and the lower solids concentrations of many chemical sludges. In some cases, phosphorus removal has shifted the economics of sludge processing favor of hauling rather than dewatering and/or incineration. In other cases, sludge hauling simply provides an interim solution to problems although it is not necessarily the most cost-effective alternative.

Cropland application of chemical sludges has raised many serious questions with regard to the transmittal of contaminants to plants and animals. The use of lime, aluminum salts, or iron salts for phosphorus removal precipitates most of the metallic cations contained in the wastewater as well as the bulk of the phosphorus. These constituents, as well as the iron, aluminum, or calcium of the precipitating chemicals,are concentrated in the sludge. Chemical sludges therefore contain nutrients and other elements which can be beneficial to plant growth. Lime sludges can improve low pH, low calcium, or low phosphorus soils. Chemical sewage sludges must be characterized on an individual basis to determine if possibly hazardous concentrations of heavy metals or other contaminants exist.

SECTION 11

STATE-OF-THE-ART APPRAISAL

It is evident that the addition of chemical sludges to the regular sewage treatment plant primary and secondary sludges often has a significant impact upon subsequent sludge handling. The volume and mass of the sludge increase and the percentage of volatile solids decreases. Except in the case of lime sludges, thickening and dewatering efficiencies are often adversely affected. Chemical conditioning requirements change. Where incineration is used, an increased need for supplemental fuel is reported. The extent and seriousness of these and other effects vary greatly between treatment plants. Each treatment plant is unique, and neither the problems nor their solutions are universal. However, certain generalizations can be made from the data obtained during this investigation.

Of the three chemicals normally considered for phosphorus removal -- lime, iron salts, or aluminum salts -- iron salts generally appear to have the least overall adverse effect upon subsequent sludge handling. This conclusion is based primarily upon two factors:

1. The addition of lime generates a much greater mass of sludge than does the addition of iron or aluminum salts.
2. The chemical sludge generated by the addition of aluminum salts is usually more difficult to thicken and/or dewater than the sludge generated by the addition of iron salts.

Obviously, for many plants these two advantages are overridden by other considerations or else iron salts would be routinely used by all plants. Other considerations include phosphorus removal efficiency, geographical variations in chemical costs and the corrosiveness of iron salts.

The next decision to be made is where in the sewage treatment chain to apply the phosphorus removal chemical. For a typical activated sludge treatment plant, there appears to be some advantage to adding iron or aluminum salts just ahead of the aeration tank or directly to the mixed liquor, where good mixing is achieved prior to discharge to the secondary clarifier. When using a ferrous iron salt, unless it is accompanied by a base, removal of phosphorus takes place at the secondary treatment stage if the chemical is added before the primary clarifier.

If lime is the chemical used, it is added to the primary treatment step or occasionally to a special tertiary treatment process. Lime is never added to the secondary biological process.

Several older plants which have inadequate volume capacity for sludge handling have found that pumping chemical-laden waste activated sludge to the primary clarifier influent for settling with the primary sludge reduced the volume of combined sludge to be treated. At these plants, sludge handling considerations have either been judged to outweigh the problem of deterioration of primary effluent quality, or an increase in secondary clarifier efficiency has counteracted the problem.

It has become common practice to thicken sludge in a gravity thickener prior to further sludge treatment. In virtually all cases, chemical sludges containing iron salts thicken much better than sludge containing aluminum salts. Best results with either iron or aluminum sludges are obtained with the addition of a polymer (dosage range 0.5 to 1.0 mg/ℓ). It has also been found that the lower the dosage of iron or aluminum salt used the easier the resulting sludge is to thicken. For this reason, and to save chemical costs, it is recommended that chemical feed equipment be automatically controlled to prevent overdosage of more chemical than required to achieve the phosphorus reduction needed.

If it is necessary to thicken secondary biological sludge separately, experience indicates that air flotation thickening is superior to gravity thickening. Again, the addition of polymers substantially improves performance.

Centrifuge dewatering of primary or combined chemical sludges is greatly enhanced by polymer addition. Sludges containing iron salts dewater much better than sludges containing aluminum salts. Lime sludges dewater very well.

Anaerobic and aerobic sludge digestion is reported to be essentially uninhibited by the addition of chemical sludges, there being no toxic effects from the presence of the chemical precipitates or the pH of the sludge. However, there is frequently the need for more volume to handle the increased sludge mass while maintaining proper retention time. When iron and aluminum sludges are added to anaerobic digesters, there is also commonly an adverse effect on supernatant quality and digested sludge solids concentration because of poor solids-liquid separation.

Vacuum filtration of chemical sludges presented problems at a number of plants due to increased solids mass, sludge volume, and/or poorer sludge dewatering characteristics. Experimentation with chemical conditioning, e.g., polymer dosages, lime addition, etc., generally led to improved vacuum filter performance. In addition, changes in filter media were reported helpful. The City of Milwaukee, Wisconsin, found in pilot tests that top feed vacuum filtration of iron sludges was more effective than conventional bottom feed filters. As was the case with centrifuges, iron sludges are generally reported easier to dewater than aluminum sludges.

Thermal conditioning of chemical sludges is generally reported successful prior to vacuum filtration or centrifugation. Sludge cakes of 35 percent TS and above are routinely achieved. Potential negative aspects are similar to those for non-chemical sludges: sidestreams have high dissolved organic strength, and operation and maintenance costs are high. One plant reported excessive corrosion and erosion of the thermal conditioning unit components, but it is not known if the problem was aggravated by the chemical component of the sludge.

Because of the impacts of phosphorus removal on sludge production and treatment, plants are now hauling more sludge in liquid and cake form to land disposal sites than before. Hauling sludge as a liquid or cake rather than dewatering or incinerating has been a common solution to many of the difficulties experienced by plants in dewatering and incinerating chemical sludges. In some cases, phosphorus removal has shifted the economics of sludge processing in favor of hauling rather than dewatering and/or incineration. Especially in the case of lime sludges, it is being found that land application of cake is preferable to lagoon storage or incineration. In other cases, hauling simply provides an interim solution to problems, although it is not necessarily the most cost effective alternative. It is frequently relied upon as an interim solution by plants which have inadequate capacity to handle the additional sludge generated by phosphorus removal with existing facilities.

In view of the large amounts of chemical sludges being applied to land, it is important that the negative or beneficial effects on plants and animals be considered. Chemical sludges contain nutrients and other elements which are beneficial to plant growth. Lime sludges can improve low pH, low calcium, or low phosphorus soils. Chemical sewage sludges must be characterized on an individual basis to determine if possibly hazardous concentrations of heavy metals or other contaminants exist.

TABLE 31. BIBLIOGRAPHY INFORMATION MATRIX

	Orientation		Sludge Type			Sludge Treatment/Disposal Steps						Costs	Sludge Characteristics
Bibliography Number	I Level of Investigation	II Audience	III Chemical Precipitant	IV Point of Addition	V Sludges Processed	VI Thickening/ Blending	VII Stabilization/ Reduction	VIII Conditioning/ Stabilization	IX Dewatering	X Heat Drying/ Incineration	XI Final Disposal	XII Cost Coverage	XIII Sludge Generation and Characteristics
1	F	E	I,L	P	5		3		4		4,7		C,R
2	F	E	A	P		1,2	2		3,4				R
3	A	E,M	A,I	A,P	1		2	2	4				R
4	P	E	A	P	7				10			C,O	
5	P	E	I,L						4	6			C,R
6	A,R	E	L				3						C
7	R	E					2,3						
8	R	E	A,I,L				3		4		1		C
9	F	E	A	A	2,7		3						C,R
10	P	E	A,I,L	P	7	1	4			7			C,R
11	P	E	A,I,L			1			4				
12	L	E	L	P	1	1			4				
13	L,P	E	A,I			1,2			1,3,4				
14	L	E,S	A,I,L								1		C
15	L	E,S	A,I,L								1		C
16	A	E	A,I	P,S	2				2		1,2		
17	L	E	A	A	7			1	4		8	C,O	
18	A	E,M	L	A	7,9	1			3	6		C,O	R
19	R	E	A,I,L				3		*	*			R
20	A	E	A	S			3		2				R

(continued)

TABLE 31 (continued)

	Orientation		Sludge Type			Sludge Treatment/Disposal Steps						Costs	Sludge Characteristics
Bibliography Number	I Level of Investigation	II Audience	III Chemical Precipitant	IV Point of Addition	V Sludges Processed	VI Thickening/ Blending	VII Stabilization/ Reduction	VIII Conditioning/ Stabilization	IX Dewatering	X Heat Drying/ Incineration	XI Final Disposal	XII Cost Coverage	XIII Sludge Generation and Characteristics
21	R	E	L		7	1				7			
22	A	E	I	P	5,7		3		4		4		C,R
23	L,P	E	L	A	7						1		C
24	L	E	A	S	5,7		2						
25	R	E	A,I,L						*				
26	R	E	A,I,L			1			3,4				
27	F,L,R	E	I				2		9				
28	L	E	A,I				2						C
29	A,L	E	A	P	1		3						C
30	A	E	I	P	5		3						C,R
31	A	E	L	A,P		1			1,3,4	6			
32	L,P	E,S	A	S	2		2,3						
33	L	E	L				2		4				
34	L	E	A,I,L				4						C
35	A	E	P				2		2				
36	R	E,S	A,I,L				3						C
37	A	E	I	P	1		2,3						
38	R	E	A,I,L						3				
39	L	E,S	A,I	P	5						1		C
40	R	E	A,I,L				3						

(continued)

TABLE 31 (continued)

Bibliography Number	Orientation		Sludge Type			Sludge Treatment/Disposal Steps						Costs	Sludge Characteristics
	I Level of Investigation	II Audience	III Chemical Precipitant	IV Point of Addition	V Sludges Processed	VI Thickening/ Blending	VII Stabilization/ Reduction	VIII Conditioning/ Stabilization	IX Dewatering	X Heat Drying/ Incineration	XI Final Disposal	XII Cost Coverage	XIII Sludge Generation and Characteristics
41	R	E	A,I,L				2						
42	L	E	I	P,S	1,5	1	3						
43	A	E	L	P	1	1		1	3	7	4	C,0	C,R
44	R	E	A,I,L				3						C
45	R	E	L	P		*	3		2,3,4	6			
46	L	E	I						4				
47	L,P	E	L	A		1			4				
48	L	E	A,I,L			1		1	10				
49	R	E,M	L								1,8	C,0	C
50	F	E	A		1,3	2	3				1		
51	A	E	A	S	3		3						C,R
52	A	E	A,I,L	A,P					4			C,0	R
53	L	S	A,I										
54	R	E	A,P						3				
55	R	E,M	I,L		2		3	1,2,3	3,4	4			
56	L	E	L										C
57	R	E	A,L,I			1	4	1,3	1,6	1,3,6,7			C,0
58	L	S									1		C
59	R	E	A,I,L								8		C
60	L	S	A,I,L				3	1	4		8		C

(continued)

TABLE 31 (continued)

	Orientation		Sludge Type			Sludge Treatment/Disposal Steps						Costs	Sludge Characteristics
Bibliography Number	I Level of Investigation	II Audience	III Chemical Precipitant	IV Point of Addition	V Sludges Processed	VI Thickening/ Blending	VII Stabilization/ Reduction	VIII Conditioning/ Stabilization	IX Dewatering	X Heat Drying/ Incineration	XI Final Disposal	XII Cost Coverage	XIII Sludge Generation and Characteristics
61	R	E	A,I,L	P	8			3	4	3,6,7		C,0	
62	L	E,S	I										C
63	R	E	A,I,L				2,3						
64	F	E	I		1		3		8				C
65	F	E	L	P	5				3		1		
67	L,P	E	A,I,L						3,4				C
68	F	E	I	P	1		2,3		2				
69	A	E	A	A					2		1		C
70	R	E,M	A,I,L			*			*	*			R
71	L	E	I,L								8	C,0	
72	R	E	A,L	P,S	1,5		3		3,4		1,4	C,0	
73	A,R	E	A,I,L		1		3	1	3,4		1,4		
74	L	S	A								1		C
75	F	S	A,I,L								1		
76	L	S	A,I,L								1		
77	P,F	E	A,I			1	2,3	1,3	3,4				C,R
78	A	E	I	S	1				2			C,0	
79	R	E	A,I,L			1,2	2,3,4		3,4	3,6,7	1		C,R
80	R	E,M	A,I,L										

TABLE 32. KEY TO BIBLIOGRAPHY INFORMATION MATRIX

I Level of investigation:

- (L) Laboratory
- (P) Pilot
- (F) Full-scale demonstration
- (A) Actual operation
- (R) Review

II Audience:

- (M) Management
- (E) Engineering
- (S) Scientific

III Chemical precipitant:

- (A) Aluminum salt
- (L) Lime
- (I) Iron salt
- (P) Polymer

IV Point of addition:

- (P) Pre-primary
- (S) Pre-secondary or secondary
- (A) After secondary

V Sludges processed:

- (1) Raw primary + WAS
- (2) Digested primary + WAS
- (3) Raw primary + TF
- (4) Digested primary
- (5) Raw primary
- (6) Digested primary
- (7) Separate chemical
- (8) Raw secondary
- (9) Digested secondary

VI Thickening/blending:

- (1) Gravity thickener
- (2) Flotation thickener
- (3) Stirred thickener
- (4) Centrifuge

VII Stabilization/reduction:

- (1) Composting
- (2) Aerobic digestion
- (3) Anaerobic digestion
- (4) Wet air oxidation
- (5) Pyrolysis
- (6) Chemical stabilization

VIII Conditioning/stabilization:

- (1) Chemical conditioning
- (2) Elutriation
- (3) Thermal conditioning
- (4) Radiation treatment
- (5) Freezing

IX Dewatering:

- (1) Pressure filter
- (2) Air drying beds
- (3) Centrifuge
- (4) Vacuum filter
- (5) Horizontal moving screen concentrator
- (6) Belt filter press
- (7) Cylindrical rotating gravity filter
- (8) Capillary suction
- (9) Lagoon
- (10) Moving bed sand filter

(continued)

TABLE 32 (continued)

X Heat drying/incineration:

- (1) Flash or spray dryer
- (2) Tray dryer
- (3) Rotary kiln furnace
- (4) Rotary kiln dryer
- (5) Multiple hearth dryer
- (6) Multiple hearth incinerator
- (7) Fluidized bed incinerator

XI Final disposal:

- (1) Agricultural fields
- (2) Land reclamation
- (3) Power generation
- (4) Sanitary landfill
- (5) Ocean disposal
- (6) Private or authority-owned dumpsite
- (7) Commercial or noncommerical soil conditioner
- (8) Resource recovery

XII Cost coverage:

- (C) Capital costs
- (O) Operating costs

XIII Sludge characteristics:

- (R) Generation rates
- (C) Chemical/physical characteristics

* Article covers information category generally

BIBLIOGRAPHY

1. Alvord, E. T., et al., "Phosphorus Removal by Ferrous Iron and Lime," 11010 EGO 01/71, U.S. Environmental Protection Agency, January 1971.

2. Baillod, C. R., Cressey, G. M., and Beaupre, R. T., "Influence of Phosphorus Removal on Solids Budget," JWPCF, 49 (1): 131-145, 1977.

3. Baldock, E. H., "Metropolitan Toronto's Experience in Phosphate Removal," Presented at Western Canada Water and Sewage Conference, September 21, 1973, Toronto.

4. Bell, G. R., Libby, D. V., and Lordi, D. T., "Phosphorus Removal Using Chemical Coagulation and a Continuous Countercurrent Filtration Process, 17010EDO 06/70, Federal Water Quality Administration, U.S. Dept. of the Interior, June 1970.

5. Bennett, S. M., and Bishop, D. F., "Solids Handling and Reuse of Lime Sludge," Contract No. 14-12-818, U.S. Environmental Protection Agency.

6. Black, S. A., "Anaerobic Digestion of Lime Sewage Sludge," Research Report No. 50, Canada-Ontario Agreement on Great Lakes Water Quality, Environment Canada, Ottawa, 1976.

7. Boyko, B. I., "Aerobic and Anaerobic Sludge Digestion," In: Proceedings of the Sludge Handling and Disposal Seminar, September 18-19, 1974, Toronto, pp. 46-62.

8. Boyko, B. I., and Rupke, J. W. G., "Phosphorus Removal within Existing Wastewater Treatment Facilities," Research Report No. 44, Canada-Ontario Agreement on Great Lakes Water Quality, Environment Canada, Ottawa, 1976.

9. Brandt, H. T., and Kuhns, R. E., "Apollo County Park Wastewater Reclamation Project, Antelope Valley, California," EPA-600/2-76-022, U.S. Environmental Protection Agency, March 1976.

10. Burns, D. E., and Shell, G. L., "Physical Chemical Treatment of a Municipal Wastewater Using Powdered Carbon." EPA-R2-73-264, U.S. Environmental Protection Agency, May 1972.

11. Burns, D. E., and Shell, G. L., "Physical-Chemical Treatment of a Municipal Wastewater Using Powdered Carbon," EPA-R2-73-264, U.S. Environmental Protection Agency, August 1973.

12. Buzzel, J. C., Jr., and Sawyer, C. N., "Removal of Algal Nutrients from Raw Wastewater with Lime," JWPCF, 39 (10): R16-R24, 1961.

13. Campbell, H., and LeClair, B. P., "Effects of Control Variables and Sludge Characteristics on the Performance of Dewatering and Thickening Devices," In: Proceedings of the Sludge Handling and Disposal Seminar, September 18-19, 1974, Toronto, pp. 280-314.

14. Chawla, V. K., Stephenson, J. P., and Liu, D., "Biochemical Characteristics of Digested Chemical Sewage Sludges," In: Proceedings of the Sludge Handling and Disposal Seminar, September 18-19, 1974, Toronto, pp. 63-94.

15. Chawla, V. K., Bryant, D. N., and Liu, D., "Disposal of Chemical Sewage Sludges on Land and Their Effects on Plants, Leachate and Soil Systems," In: Proceedings of the Sludge Handling and Disposal Seminar, September 18-19, 1974, Toronto, pp. 207-233.

16. Connell, C. H., "Phosphorus Removal and Disposal from Municipal Wastewater," 17010 DYB02/71, U.S. Environmental Protection Agency, February 1971.

17. Cornwell, D. A., and Zoltek, J. Jr., "Recycling of Alum Used for Phosphorus Removal in Domestic Wastewater Treatment," JWPCF, 49(4): 600-612, 1977.

18. Culp, R. L., Evans, D. R., and Wilson, J. C., "Advanced Wastewater Treatment as Practiced at South Tahoe," EPA-WQO-17010-ELQ-08/71, U.S. Environmental Protection Agency, August 1971.

19. Culp, G., Suhr, L. G., and Evans, D. R., "Physical-Chemical Wastewater Treatment Plant Design," EPA Technology Transfer Seminar Publication, August 1973.

20. Derrington, R. E., Stevens, D. H., and Laughlin, J. E., "Enhancing Trickling Filter Plant Performance by Chemical Precipitation," EPA-670/2-73-060, U.S. Environmental Protection Agency, August 1973.

BIBLIOGRAPHY (Continued)

21. Dorr-Oliver, Inc., "Phosphate Extraction Process," Stamford, Conn., 1968.

22. Dow Chemical Company, "Application of Chemical Precipitation Phosphorus Removal at the Cleveland Westerly Wastewater Treatment Plant," Prepared for the City of Cleveland, Ohio, April 1970.

23. Dunseth, M. G., et al., "Ultimate Disposal of Phosphate from Waste Water by Recovery as Fertilizer," 17070ESJ01/70, Federal Water Quality Administration, U.S. Dept. of the Interior, January 1970.

24. Eikum, A. S., Carlson, D. A., and Lundar, A., "Phosphorus Release during Storage of Aerobically Digested Sludge," JWPCF, 47(2): 330-337, 1975.

25. Farrell, J. B., "Design Information on Dewatering Properties of Wastewater Sludges," In: Proceedings of the Sludge Handling and Disposal Seminar, September 18-19, 1974, Toronto, pp. 269-279.

26. Farrell, J. B., "Interim Report of Task Force on Phosphate Removal Sludges," EPA-670/2-75-013, U.S. Environmental Protection Agency, 1975.

27. Fowlie, P. J. A., and Shannon, E. E., "Utilization of Industrial Wastes and Waste By-Products for Phosphorus Removal: An Inventory and Assessment," Research Report No. 6, Canada-Ontario Agreement on Great Lakes Water Quality, Environment Canada, Ottawa, June 1973.

28. Ganczarczyk, J., and Hamoda, M.F.D., "Aerobic Digestion of Organic Sludges Containing Inorganic Phosphorus Precipitates, Phase 1," Research Report No. 3, Canada-Ontario Agreement on Great Lakes Water Quality, Environment Canada, Ottawa, June 1973.

29. Gray, I. M., "Phosphorus Removal Study at Barrie WPCP," Research Branch, Ontario Ministry of the Environment, Toronto, 1972.

30. Gray, I. M., "Phosphorus Removal Study at the Sarnia WPCP," Research Report No. 14, Canada-Ontario Agreement on Great Lakes Water Quality, Environment Canada, Ottawa, 1972.

BIBLIOGRAPHY (Continued)

31. Greenlund, T. W., and Gaines, F. R., "Hatfield Township, Pennsylvania Advanced Waste Treatment Plant," Tracy Engineers, Inc., Camp Hill, Pennsylvania, 1977.

32. Grigoropoulos, S. G., Vedder, R. C., and Max, D. W., "Fate of Aluminum-Precipitated Phosphorus in Activated Sludge and Anaerobic Digestion," JWPCF, 43(12): 2366-2382, 1971.

33. Hamoda, M. F., and Ganczarczyk, J., "Aerobic Digestion of Sludges Precipitated from Wastewater by Lime Addition," JWPCF, 49(3): 375-387, 1977.

34. Hudgins, R. R., and Silveston, P. L., "Wet Air Oxidation of Chemical Sludges," Research Report No. 12, Canada-Ontario Agreement on Great Lakes Water Quality, Environmental Canada, Ottawa, March 1973.

35. Jacke, R., "Polymer Cuts Disposal Costs," Water and Sewage Works, 123(6): 99-100, 1976.

36. Jenkins, D., Ferguson, J. F., and Menar, A. B., "Chemical Processes for Phosphate Removal," Water Research, 5: 369-389, 1971.

37. Johnson, E. L., Beeghly, J. H., and Wukasch, R. F., "Phosphorus Removal at Benton Harbor-St. Joseph, Michigan," Dow Chemical Company, Midland, Michigan, 1969.

38. Keith, F. W., Jr., "Centrifuges--Types and Applications," In: Proceedings of the Sludge Handling and Disposal Seminar, September 18-19, 1974, Toronto, pp. 351-368.

39. Kirkham, M. B., and Dotson, G. K., "Growth of Barley Irrigated with Wastewater Sludge Containing Phosphate Precipitants," In: Proceedings of the National Conference on Municipal Sludge Management, June 11-13, 1974, pp. 97-106.

40. Knight, C. H., Mondoux, R. G., and Hambley, B., "Thickening and Dewatering Sludges Produced in Phosphate Removal," Paper presented at Phosphorus Removal Design Seminar, May 28-29, 1973, Toronto.

BIBLIOGRAPHY (Continued)

41. Koers, D. A., "Aerobic Digestion of Wastewater Sludges," Paper presented at Technology Transfer Seminar on Sludge Handling Disposal, February 16-19, 1977, Calgary, Alberta.

42. Malhotra, S. K., Parrillo, T. P., and Hartenstein, A. G., "Anaerobic Digestion of Sludges Containing Iron Phosphates," Journal of the Sanitary Engineering Division, ASCE, 97(SA5): 629-646, October 1971.

43. Martin, L., and Nardozzi, A. D., "Operational Considerations Associated with Chemical and Biological Sludge Generated by Lime Precipitation," Paper presented at the National WPCF Conference, October 1976, Minneapolis, Minnesota.

44. Mignone, N. A., "Anaerobic Digester Supernatant Does Not Have to be a Problem," Water and Sewage Works, 123(12): 57-59, 1976.

45. Minton, G. R., and Carlson, D. A., "Primary Sludges Produced by the Addition of Lime to Raw Waste Water," Water Research, 7: 1821-1847, 1973.

46. Motamedi, M., "Dewatering of Ferric Chloride Coagulation Sludge," Water Research, 9: 861-864, 1975.

47. Mulbarger, M. C., et al., "Lime Clarification, Recovery, Reuse, and Sludge Dewatering Characteristics," JWPCF, 41(12): 2070-2085, 1969.

48. Novak, J. T., and Montgomery, G. E., "Chemical Sludge Dewatering on Sand Beds," Journal of the Environmental Engineering Division, ASCE, 101(EE1): 1-14, February 1975.

49. Opferkuch, R. E., Ctvrtnicek, T., and Mehta, S. M., "Study of Utilization and Disposal of Lime Sludges Containing Phosphates," EPA-R2-73-282, U.S. Environmental Protection Agency, June 1973.

50. O'Shaughnessy, J. C., et al., "Digestion and Dewatering of Phosphorus-Enriched Sludges," JWPCF, 46(8): 1914-1926, 1974.

BIBLIOGRAPHY (Continued)

51. O'Shaughnessy, J. C., et al., "Soluble Phosphorus Removal in the Activated Sludge Process. Part II: Sludge Digestion Study," EPA-17010-EIP-10/71, U.S. Environmental Protection Agency, October 1971.

52. Proceedings of the Technical Seminar on Physical-Chemical Treatment, Ontario Ministry of Health Laboratories, March 9, 1972.

53. Recht, H. L., and Ghassemi, M., "Kinetics and Mechanism of Precipitation and Nature of the Precipitate Obtained in Phosphate Removal from Wastewater Using Aluminum (III) and Iron (III) Salts," 17010EKI 04/70, Federal Water Quality Administration, U.S. Dept. of the Interior, April 1970.

54. Roe, I. P., "Sludge Dewatering Using the Kruger Centrifuge," In: Proceedings of the Sludge Handling and Disposal Seminar, September 18-19, 1974, Toronto, pp. 369-384.

55. Salib, W. A., "Sludge Handling and Disposal Practices in Metropolitan Toronto," In: Proceedings of the Sludge Handling and Disposal Seminar, September 18-19, 1974, Toronto, pp. 452-465.

56. Schmid, L. A., and McKinney, B. E., "Phosphate Removal by a Lime-Biological Treatment Scheme," JWPCF, 41(7): 1259-1276, 1969.

57. Schroeder, W. H., "Principles and Practices of Sludge Incineration," Paper presented at the Technology Transfer Seminar on Sludge Handling Disposal, February 1977, Calgary, Alberta.

58. Scott, D. S., and Horlings,H., "Removal of Phosphates and Metals from Sewage Sludge," Research Report No. 28, Canada-Ontario Agreement on Great Lakes Water Quality, Environment Canada, Ottawa, 1973.

59. Scott, D. S., and Horlings, H., "Removal of Phosphates and Metals from Sewage Sludges," In: Proceedings of the Sludge Handling and Disposal Seminar, September 18-19, 1974, Toronto, pp. 413-443.

60. Scott, D. S., and Horlings, H., "Removal of Phosphates and Metals from Sewage Sludges," Environmental Science and Technology, 9(9): 849-855, 1975.

BIBLIOGRAPHY (Continued)

61. Shannon, E. E., Plummer, D., and Fowlie, P.J.A., "Aspects of Incinerating Chemical Sludges," In: Proceedings of the Sludge Handling and Disposal Seminar, September 18-19, 1974, Toronto, pp. 391-412.

62. Singer, P. C., "Anaerobic Control of Phosphate by Ferrous Iron," JWPCF, 44(4): 663-669, 1972.

63. Smart, J., "Anaerobic Sludge Digestion Processes," Paper presented at Technology Transfer Seminar on Sludge Handling Disposal, February 1977, Calgary, Alberta.

64. Smart, J., and Boyko, B. I., "Full Scale Studies on the Thermophilic Anaerobic Digestion Process," Research Report No. 59, Canada-Ontario Agreement on Great Lakes Water Quality, Environment Canada, Ottawa, 1977.

65. Smith, A. G., "Centrifuge Dewatering of Lime Treated Sewage Sludge," Paper No. W2030, Research Branch, Ontario Ministry of the Environment, Toronto, 1972.

66. Stickney, R., and LeClair, B. P., "The Use of Physico-chemical Sludge Characteristics and Bench Dewatering Tests in Predicting the Efficiency of Thickening and Dewatering Processes," In: Proceedings of the Sludge Handling and Disposal Seminar, September 18-19, 1974, Toronto, pp. 315-350.

67. "Studies on Removal of Phosphates and Related Removal of Suspended Matter and Biochemical Oxygen Demand," May-October 1967, Lake Odessa, Michigan, Wastewater Section, Division of Engineering, Michigan Dept. of Public Health.

68. Tofflemire, T. J., et al., "Tertiary Treatment for Phosphorus Removal by Alum Addition at Richfield Springs, New York," Research Paper No. 39, Environmental Quality Research Unit, New York State Dept. of Environmental Conservation, March 1976.

69. U.S. Environmental Protection Technology Transfer, "Physical-Chemical Wastewater Treatment Plan Design," EPA Technology Transfer Seminar Publication, 1973.

70. University of Guelph, Ontario, "Land Disposal of Sewage Sludge. Vol. I," Research Report No. 16, Canada-Ontario Agreement on Great Lakes Water Quality, Environment Canada, Ottawa, 1973.

BIBLIOGRAPHY (Continued)

71. Van Fleet, G. L., Barr, J. R., and Harris, A. J., "Treatment and Disposal of Chemical Phosphate Sludge in Ontario," JWPCF, 46(3): 582-587, 1974.

72. Van Fleet, G. L., Barr, J. R., and Harris, A. J., "Treatment and Disposal of Chemical Phosphate Sludge in Ontario," Presented by J. R. Barr at the Annual Conference of the Water Pollution Control Federation, October 1972, Atlanta, Georgia.

73. Van Loon, J. C., "Heavy Metals in Agricultural Lands Receiving Chemical Sewage Sludges," Research Report No. 9, Canada-Ontario Agreement on Great Lakes Water Quality, Environment Canada, Ottawa, March 1973.

74. Van Loon, J. C., "Heavy Metals in Agricultural Lands Receiving Chemical Sewage Sludges. Vol. II," Research Report No. 25, Canada-Ontario Agreement on Great Lakes Water Quality, Environment Canada, Ottawa, 1975.

75. Van Loon, J. C., "Heavy Metals in Agricultural Lands Receiving Chemical Sewage Sludges. Vol. III," Research Report No. 30, Canada-Ontario Agreement on Great Lakes Water Quality, Environment Canada, Ottawa.

76. Villiers, R. V., "Thickening and Dewatering Phosphorus-Laden Chemical Sludges, Field Operation--1974," U.S. Environmental Protection Agency, 1974.

77. Williams, T. C., "Phosphorus is Removed at Low Cost," Water and Wastes Engineering, 13(11): 52-63, 1976.

78. Wood, G. M., "Land Application of Processed Organic Wastes," In: Proceedings of the Phosphorus Removal Design Seminar Conference, May 28-29, 1973, Environment Canada, Toronto.

79. Zaloum, R., "Activated Sludge Characterization and Settling," In: Proceedings of Phosphorus Removal Design Seminar, May 28-29, 1973, Environment Canada, Toronto.

OMB #158-S-77002
February 1978

APPENDIX A

PHOSPHORUS-LADEN SLUDGE MANAGEMENT QUESTIONNAIRE

When completed, mail to:

SCS Engineers
4014 Long Beach Boulevard
Long Beach, California 90807
(213) 426-9544

I. GENERAL INFORMATION

1. Full plant name:________
2. Address:________
3. Phone number:________
4. Name of person completing questionnaire:________
5. Title:________
6. Name of alternate contact for technical information:________
7. Title:________
8. Approximate time when (a) primary plant was built:________
   (b) secondary treatment was installed:________
   (c) phosphorus removal was begun:________
9. Has the plant ever participated directly in any EPA, university, or other study pertinent to phosphorus-laden sludge management?
   Yes________ No________

II. RAW SEWAGE INFLUENT INFORMATION (Recent monthly averages)

1. Volume:________mgd
2. pH:________ BOD_5:________mg/ℓ COD:________mg/ℓ SS:________mg/ℓ
   Total-P:________mg/ℓ VSS________mg/ℓ
3. Estimated percentage of industrial wastes________%
4. Do you get sludge from the water purification plant? Yes________
   No________ Don't know________
5. Storm and sanitary sewers combined________ or separate________ (check one)

III. PLANT DESIGN INFORMATION (Please attach flow diagram if available).

1. Primary treatment (circle below)
   - Screening
   - Aerated grit chamber
   - Non-aerated grit chamber
   - Gravity settling
   - Flocculation chamber

   Others (please explain)______________________________

   ____
2. Secondary treatment (circle below)
   - Conventional activated sludge
   - Trickling filter
   - Final clarifier
   - Modified activated sludge:
     - Step-aeration
     - Contact stabilization
     - Extended aeration
     - Complete mix
     - Pure oxygen

   Others (please explain)______________________________

   ____
3. Sludge treatment/disposal - Circle boxes on Figure 1 on the next page and connect with arrows to show your sludge treatment/disposal chain. Indicate any points of chemical addition as well.

IV. PHOSPHORUS REMOVAL INFORMATION (Give recent monthly averages)

1. Effluent total-P:_____mg/ℓ
2. Chemicals added and chemical dosage -

a. Chemicals added to remove phosphorus: (circle below)	b. Percent of chemical in solution added:*		c. Average amounts added:*
Lime	% $Ca(OH)_2$	____	__________gpd
Ferric chloride	% $FeCl_3$	____	__________gpd
Ferrous sulfate	% $FeSO_4$	____	__________gpd
Alum(aluminum sulfate)	% $Al_2(SO_4)_3$	____	__________gpd
Sodium aluminate	% $NaAlO_2$	____	__________gpd
Polymer			__________gpd
Other(name below)	% __________	____	__________gpd

____

*Explanation of parts b and c: We assume you add a solution containing a certain percentage of actual chemical to the wastewater. Give that percentage in part b. Then tell how much of that solution you add in part c. Use other units such as gal/mil gal of sewage rather than gpd if you wish, but please indicate units. If you wish to respond to this section in another way, by giving the dosage of dry chemical in mg/ℓ or lb/day, for example, write the appropriate information below:

____

____

____

3. Indicate the sources of the chemicals by circling below and writing the name of the chemical next to its source:
   Chemical manufacturer________________________
   Industrial waste product______________________
   Drinking water purification sludge______________
   Other (specify)________________ ______________
4. Point of chemical addition: (circle below)

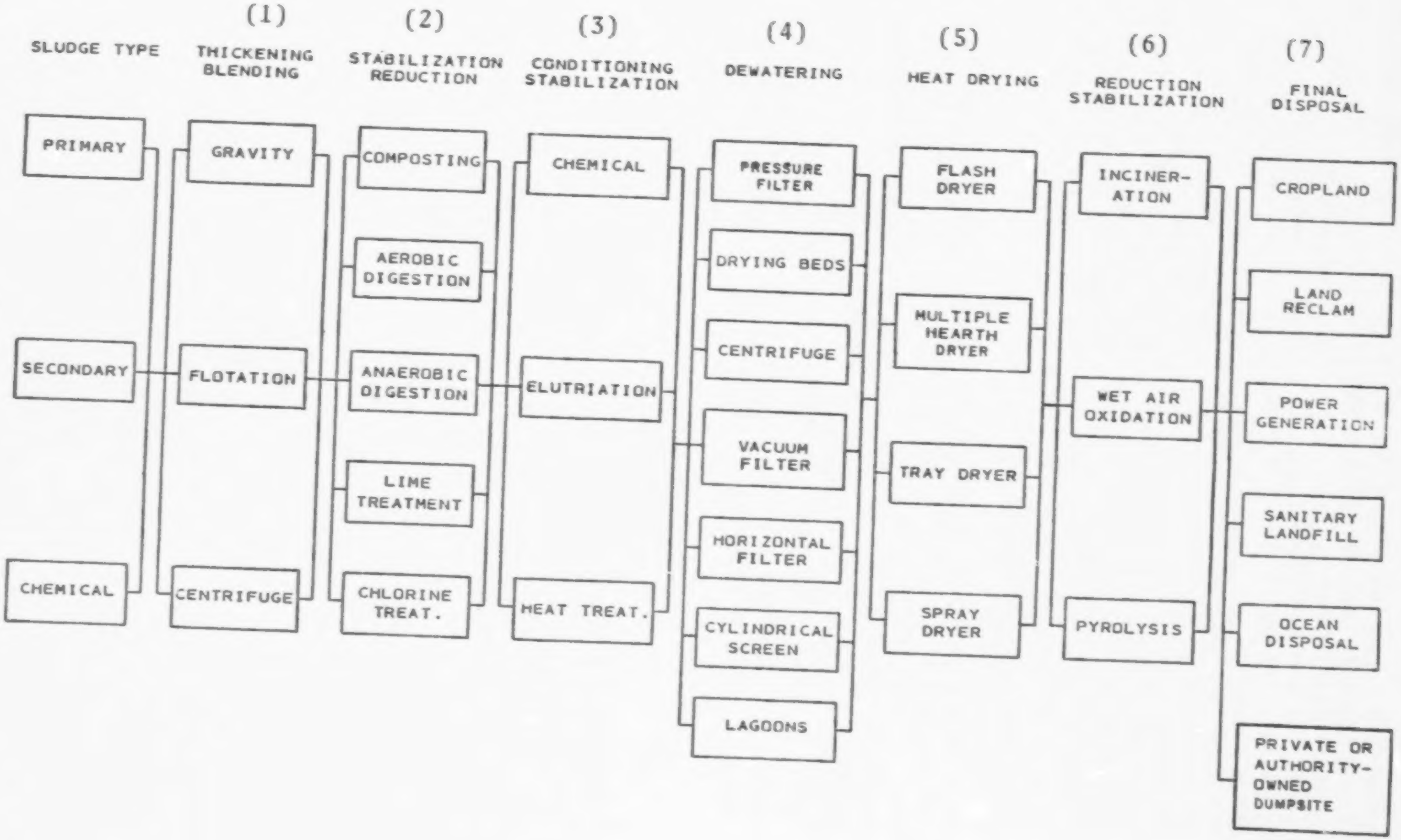
Before primary clarifier	Primary clarifier
Activated sludge tank	Trickling filter
Activated sludge effluent channel	Trickling filter effluent channel or pipe
After secondary clarifier	Secondary clarifier
Other (please explain)	

   ____

V. SLUDGE HANDLING INFORMATION

1. Does your plant combine all of its sludges --primary, waste activated or trickling filter, and chemical -- together for dewatering and disposal, etc.?
   Yes_________ No_________
2. If answer to previous question #1 is no, indicate what sludges you treat or dispose of individually: (circle below)
   Primary   Secondary   Chemical
   Combined Primary/Secondary   Other (specify)________________________
3. Chemical conditioners used for dewatering: (circle below)
   Lime   Ferric chloride   Polymer   Others (specify)
   ____

Circle applicable boxes and connect with lines to show sludge treatment/disposal chain.


SLUDGE TYPE
PRIMARY
SECONDARY
CHEMICAL
(1)
THICKENING BLENDING
GRAVITY
FLOTATION
CENTRIFUGE
(2)
STABILIZATION REDUCTION
COMPOSTING
AEROBIC DIGESTION
ANAEROBIC DIGESTION
LIME TREATMENT
CHLORINE TREAT.
(3)
CONDITIONING STABILIZATION
CHEMICAL
ELUTRIATION
HEAT TREAT.
(4)
DEWATERING
PRESSURE FILTER
DRYING BEDS
CENTRIFUGE
VACUUM FILTER
HORIZONTAL FILTER
CYLINDRICAL SCREEN
LAGOONS
(5)
HEAT DRYING
FLASH DRYER
MULTIPLE HEARTH DRYER
TRAY DRYER
SPRAY DRYER
(6)
REDUCTION STABILIZATION
INCINERATION
WET AIR OXIDATION
PYROLYSIS
(7)
FINAL DISPOSAL
CROPLAND
LAND RECLAM
POWER GENERATION
SANITARY LANDFILL
OCEAN DISPOSAL
PRIVATE OR AUTHORITY-OWNED DUMPSITE

VI. SLUDGE CHARACTERISTICS (Recent monthly averages)

1. Sludge characteristics before thickening or stabilization by digestion, heat, or chemical treatment:
   (For those who answered question V-2 above, there is room for the characteristics of more than one type of sludge.)

Type of sludge:				Substitute other units (e.g., mg/ℓ)
Volume (gpd)				( )
TS (%, dry wt)				( )
VS (% of TS)				( )
COD (mg/ℓ)				( )

2. Sidestreams generated by sludge treatment units: (circle below)
   Centrate   Digester supernatant
   Filtrate   Thickener supernatant   Other(specify)__________
3. Sidestream characteristics:

Type of sidestream:				Substitute other units (e.g., mg/ℓ, TS, etc.)
Volume (gpd)				( )
SS (%, dry wt)				( )
VSS(% of SS)				( )
COD(mg/ℓ)				( )
BOD$_5$(mg/ℓ)				( )

4. Final TS content of treated liquid sludge before drying, dewatering, or ultimate disposal: ____% TS, dry wt.

VII. PHOSPHORUS REMOVAL IMPACT ON PLANT OPERATION

1. Additional sludge generated - (Give typical monthly averages)
   a. Total sludge volume before plant had phosphorus removal: _______gpd
   b. Sludge TS before plant had phosphorus removal: % dry wt______, or mg/ℓ____
   c. Wastewater flow volume before plant had phorphorus removal: _______mgd
   d. Wastewater SS before plant had phosphorus removal:________mg/ℓ
   e. If possible, estimate how much additional sludge is generated as a result of phosphorus removal: ___________#/day dry solids
2. How have changes in sludge volume or mass and sludge characteristics affected plant operations? For instance, does the phosphorus-laden sludge require more time, energy, equipment, etc. to treat/dispose of than the sludge before P-removal did? How? Is the phosphorus-laden sludge more difficult to settle, thicken, dewater, digest, ect.? How? How did you solve any problems which resulted from phosphorus removal? Please respond to these questions in as much detail as possible. This is the most important part of the questionnaire. We appreciate your thoughtful response.

Please use other side or extra pages if needed.

3. If not covered above, please discuss any adverse effects of the phosphorus-laden sludge on anaerobic digestion?

4. Amount of extra labor, energy, equipment, and supplies needed because of changes in sludge volume or mass and sludge characteristics (please estimate)-
   a. New equipment for sludge treatment/disposal:

Equipment item	Year Added	Capital cost ($)	Estimated operating cost ($/yr)	Year

   b. Additional labor for maintenance and operation of all sludge treatment/disposal equipment:

Area where additional labor was needed	Number of additional labor hours (hr/yr)	Cost ($/hr)

c. Additional energy and fuel requirements for operation of all sludge treatment/disposal equipment:

Equipment item or area of impact	Type of energy/fuel	Number of additional KWH, MCF, etc./yr	Cost ($/KWH or MCF, etc.)

d. Additional chemical supplies for sludge treatment/disposal: (Include chemicals for conditioning, thickening, oxidation, etc.)

Chemical	Area of use of chemical	Additional lb of chemical, (cu ft oxygen),etc./day	Cost($/lb or cu ft)

e. Additional outside costs related to sludge treatment/disposal (e.g., increased sludge volume means must pay contractor more to haul it away):

Type of additional outside cost	Cost($/yr)	Year

5. Effects of P-removal- (Answer only questions which apply to your plant and for which you have available data or can estimate. You may answer part of a question).

a. Incineration or mechanical heat drying rate (#wet cake/hr*)
Increase?______ Before =______#wet cake/hr
Decrease?______ After = ______#wet cake/hr
Same?______

b. Dewatering rate (#dry sludge/hr*)
Increase?______ Before =______ #dry sludge/hr
Decrease?______ After = ______ #dry sludge/hr
Same?______

*Note parts a & b: If the size of the dewatering device, incinerator, or mechanical heat dryer was changed between the "before" and "after" periods, report the "yield" instead of the "rate" by dividing by the effective area or volume of the device, and note this change.

c. Cost of chemical conditioning ($/dry ton solids dewatered):

Increase?_______ Before = _______ $/dry ton solids
Decrease?_______ After = ________ $/dry ton solids
Same?___________

d. Dewatered cake TS content (% TS, dry wt):

Increase?_______ Before = _______%, dry wt
Decrease?_______ After = ________%, dry wt
Same?___________
(Current dewatered cake VS content (% of TS): _____%)

Thank you for your cooperation in providing us information vital to the successful completion of this research. If there are any questions, please contact Lee Hammer or Michael Swayne at SCS Engineers, phone collect: (213) 426-9544.

APPENDIX B

OUTLINE FOR COLLECTION OF FIELD DATA

I. GENERAL INFORMATION

- A. Full plant name; operating authority name
- B. Addresses
- C. Phone numbers
- D. Names and titles of personnel interviewed by SCS
- E. Approximate times when all additions and modifications to plant affecting sludge were made
- F. History of experimentation at plant with modes of operation affecting sludge
    1. Points of chemical addition
    2. Chemicals used for P-removal
    3. Chemical conditioners used
    4. Equipment used
    5. Machine and process operating parameters
    6. Inclusion of chemical sludge with other sludges
    7. Disposal method
- G. Previous studies done
- H. Design engineer name, address, and phone
- I. Equipment manufacturers' names, addresses, and phones
- J. Engineering consultant's name, address, and phone
- K. Unusual wastes accepted at plant
- L. Variations in flow
- M. Before and after periods dates and modes of operation
- N. Design flow
- O. Average flow
- P. Variations in wastewater treatment operations
- Q. Unusual problems in plant operation affecting plant performance

II. GENERAL DESCRIPTION OF WASTEWATER TREATMENT OPERATIONS

- A. Diagram of wastewater treatment facilities
    1. Treatment units and arrangement

B. Machine operating parameters

1. Grit chamber (if chemical is added here)
2. Flocculators (if chemical is added here)
3. Clarifiers
4. Activated sludge or trickling filter units
5. Channels where chemicals may be added
6. Tertiary treatment units

    a. dimensions
    b. type

III. PROCURING, STORING, MIXING, PUMPING, DISTRIBUTION, AND ADDITION OF CHEMICALS FOR PHOSPHORUS REMOVAL

A. Diagram of chemical handling operations

1. Handling facilities and arrangement

B. Description of all procedures pertaining to these operations

1. Quantities and rates
2. Types of chemicals
3. Types of equipment
4. Automated or manual handling and addition
5. Locations
6. Timing
7. Personnel

IV. DETAILED DESCRIPTION OF SLUDGE TREATMENT AND DISPOSAL OPERATION

A. Diagram of sludge treatment/disposal facilities including sidestream and sludge recirculation (treatment units and arrangement)

1. Storage (including storage in clarifiers)
2. Screening, degritting
3. Pumping
4. Blending
5. Thickening
6. Conditioning
7. Stabilizing
8. Dewatering
9. Reduction
10. Disposal
11. Sidestream treatment

B. Machine operating parameters

1. Dimensions
2. Shape

3. Type
4. Brand name
5. Mechanical variations
6. Continuous or batch flows accepted
7. Expected life of equipment
8. Etc.

C. Machine operating variables

1. Speed
2. Temperature
3. Pressure
4. Vacuum
5. Cycle time
6. Fuel consumption
7. Etc.

D. Process operating variables

1. Hydraulic flow rates in and out
2. Chemical addition rates
3. Run lengths
4. Solids flow rate
5. Solids loading
6. Hydraulic loading
7. Etc.

E. Process performance variables (excluding screening, degritting, and pumping)

1. Input and output stream characteristics

    a. total solids content
    b. sludge solids content
    c. volatile solids content
    d. BOD or COD concentration
    e. total phosphorus concentration
    f. heavy metals concentrations
    g. etc.

2. Performance characteristics

    a. yield
    b. volatile solids reduction
    c. volume reduction
    d. cake thickness
    e. etc.

F. Diagrams of hydraulic and materials balances (excluding screening, degritting, and pumping)

1. Partition of loading between outputs

        a. TS
        b. sludge solids
        c. VS
        d. BOD or COD
        e. total-P
        f. heavy metals

V. OPERATIONAL PROBLEMS ATTRIBUTABLE TO THE PRODUCTION OF CHEMICAL SLUDGE

  A. Impact on wastewater treatment operations

    1. Changes in process performance variables

      a. clarifiers
      b. activated sludge or trickling filter units

        1) input and output characteristics of all units

          a) SS concentration
          b) BOD or COD concentration
          c) VSS concentration
          d) total-P concentration
          e) heavy metals concentration
          f) viruses, etc.
          g) pH, alk., sol. org. carbon

        2) performance characteristics

          a) settleability or sludge volume index of sludge
          b) #SS produced/#BOD removed in activated sludge (or TF)
          c) #/day BOD removed in activated sludge (or TF)

    2. Changes in hydraulic and materials balances of primary and secondary treatment units

      a. wastewater flow
      b. SS
      c. BOD or COD
      d. VSS
      e. others = P, pH, alk. (all forms), hardness (all forms)

  B. Additional sludge quantities, including sidestreams

    1. From clarifiers
    2. From thickeners
    3. From conditioners

4. From stabilizers
5. From dewatering unit
6. From reduction unit
7. To disposal
8. To sidestreams

    a. additional mass of dry sludge

        1) observed mass
        2) calculated mass

    b. additional volume of liquid sludge

        1) observed volume

C. Impact on sludge treatment and disposal operations

1. Impact on facilities

    a. uninitiated changes in process operating variables

        1) flow rates
        2) loading rates
        3) retention times
        4) etc.

    b. impact on physical condition and functioning of equipment, including functioning of new equipment

        1) corrosion
        2) grit
        3) pH
        4) mixing
        5) foaming
        6) wear
        7) mechanical failure

2. Impact on performance

    a. changes in process performance variables

        1) input and output characteristics

            a) TS, VS, sludge solids, SS (sidestreams)
            b) BOD, COD
            c) etc.

        2) performance characteristics

            a) yield
            b) volume reduction

            c) volatile solids reduction
            d) etc.

        b. changes in hydraulic and materials balances

            1) partition of loading between outputs

                a) TS, VS, sludge solids
                b) BOD or COD
                c) etc.

D. Problem resolution

    1. Personnel-initiated changes in machine and process operating variables

        a. machine temperature, pressure, etc.
        b. chemical addition
        c. fuel consumption, etc.
        d. run lengths
        e. etc.

    2. New construction and equipment required

        a. type, space required, etc.
        b. installation time, safety measures, etc.

    3. New operation and maintenance activities

        a. cleaning, adjustments, breakdowns, etc.
        b. setting levels, monitoring, starting and shutting off, keeping records
        c. amount of labor involved
        d. new personnel required

E. Impact on meeting external constraints restricting sludge treatment and disposal operations

    1. Environmental constraints

        a. moisture content of landfilled sludge; heavy metal content of agricultural land-applied sludge; etc.

    2. Economic constraints

        a. budget limitations; specific chemical or fuel costs, etc.

    3. Land and materials availability constraints

        a. land for disposal; industrial by-product chemicals, etc.

4. Social-acceptability constraints

   a. putrefaction odor control; disposal site location, etc.

VI. OPERATIONAL COSTS OF SLUDGE TREATMENT/DISPOSAL

A. Operations

   1. Phosphorus removal operations

      a. procuring chemicals
      b. storing chemicals
      c. mixing chemicals
      d. pumping chemicals
      e. distribution of chemicals
      f. addition of chemicals

   2. Wastewater treatment operations

      a. flocculators and clarifiers
      b. activated sludge or trickling filter units
      c. return sludge pumping

   3. Sludge treatment/disposal operations

      a. storage (including sludge storage in clarifiers)
      b. screening, degritting
      c. pumping (including sludge and sidestream recirculation)
      d. blending
      e. thickening
      f. conditioning
      g. stabilizing
      h. dewatering
      i. reduction
      j. disposal
      k. sidestream treatment

B. Costs

   1. Operating and maintenance costs

      a. labor costs

         1) operation and supervision
         2) maintenance

      b. supply costs (costs at the site)

         1) chemicals

            a) solids
            b) liquids
            c) gases

        2) water
        3) maintenance supplies

    c. energy costs (costs at the site)

        1) natural gas
        2) electricity
        3) fuel oil

    d. outside servicing and contracting

        1) service and maintenance of equipment
        2) residual disposal

    e. overhead

2. Investment costs

    a. construction

        1) site preparation
        2) structures
        3) buildings

    b. mechanical equipment on an installed basis
    c. piping on an installed basis
    d. electrical on an installed basis
    e. instrumentation on an installed basis
    f. engineering design
    g. land

C. Cost variables

1. Items or categories of labor, supplies, energy, and services utilized

2. Unit costs of labor, supply, energy, and service items

3. Amounts of labor, supply, energy, and service items used

4. Components of outside servicing and contracting, overhead, engineering design, construction, and land costs

5. Cost break-down by components of outside servicing and contracting, overhead, engineering design, construction, and land costs

6. Description of mechanical equipment, instrumentation, piping, electrical, construction, and land items.

COST MODULE

A. Investment Costs

1. Land: value/acre in general locality $____________

   Treatment system acres ____________ $/acre ____________

   Auxiliary support acres (includes utilities, offices, roads, etc.) ____________ $/acre ____________

2. Total investment (excluding land and support buildings) and year of module installation:

   $____________ Year____________

3. Installation time (excluding design) __________ months,

   useful life ______________, salvage value (items)

   ________________________ $ ____________

4. Structures (steel, dikes, slabs): $ ____________

   ________________________ $ ____________

   ________________________ $ ____________

   ________________________ $ ____________

5. Piping and valves (type, length, size): $ ____________

   ________________________ $ ____________

   ________________________ $ ____________

   ________________________ $ ____________

6. Mechanical equipment (type, size): $ ____________

   ________________________ $ ____________

   ________________________ $ ____________

   ________________________ $ ____________

7. Electrical system: $ ____________

   ________________________ $ ____________

   ________________________ $ ____________

Cost Module (Continued)

8. Instrumentation (type): $ ____________

____________________________________ $ ____________

____________________________________ $ ____________

9. Support (enclosures, access): $ ____________

____________________________________ $ ____________

____________________________________ $ ____________

10. Engineering design cost: $ ____________

11. Legal and administrative cost: $ ____________

12. How was system financed? ____________________________

____

____

____

13. Was module individual or part of a larger installation?

____

____

____

____

B. Operation and Maintenance Costs for 19__

1. Days per year of operation __________

2. Operation assumptions (flow or activity/day) __________

3. Labor requirements:

Category	Hrs/day	$/hr
a. ________________	_______	______
b. ________________	_______	______
c. ________________	_______	______

Cost Module (Continued)

4. Energy requirements:

Electricity KWH/day ___________, $/KWH ___________

Fuel (unit)__________/day, $/unit ______________

5. Chemical costs

Type __________________ $/unit ___________ as _________

Type __________________ $/unit ___________ as _________

Type __________________ $/unit ___________ as _________

6. Other supplies (type)

____ $/unit ________

7. Maintenance (labor and parts) $/year ________

8. Overhead (administration)

Percent attributed to module __________ $/year ________

9. Residual disposal

- Transportation ______________________ $/unit ________
- Disposal ______________________ $/unit ________

10. Laboratory or analysis costs (type)

____ $/unit ________

APPENDIX C

CASE STUDIES

INTRODUCTION TO CASE STUDIES

Nine case study sites were selected which were rated relatively high on the basis of completeness of historical operating data, cooperation of plant personnel, a history of problem-solving and innovation at the plant, performance of dewatering devices, and potential for demonstrating the impacts of phosphorus removal on sludge. The plants were also chosen to represent a variety of possibilities with respect to the following factors:

- Plant size
- Plant type
- Phosphorus removal chemical(s)
- Point(s) of chemical addition
- Point of combination of chemical-laden and other sludges
- Sludge treatment and disposal methods.

Table C-1 presents a description of the nine plants with respect to these factors.

CASE STUDY C: SOUTH BEND, INDIANA

Introduction

The South Bend Wastewater Treatment Plant adopted tertiary treatment with lime in 1974 in anticipation of stricter Indiana effluent regulations covering BOD, SS, and dissolved oxygen. The existing secondary treatment plant was found incapable of meeting the anticipated effluent requirements without considerable enlargement of plant capacity. An upflow clarifier system was installed to provide the tertiary treatment because it permits effective polishing of the effluent to remove BOD and SS as well as phosphorus. If phosphorus removal alone had been the objective, tertiary treatment with lime probably would not have been chosen. While the tertiary facilities were under construction, a study by Tenech, a wastewater engineering consulting firm for the city of South Bend, indicated that lime should not be used because of the high cost and the huge amounts of sludge which would be generated - an estimated 907 t (100 tons)/day.

TABLE C-1. DESCRIPTION OF CASE STUDY SITES ACCORDING TO PLANT SELECTION FACTORS

	South Bend	Sheboygan	Coldwater	Midland	Port Huron	Pontiac	Lakewood	Mentor	Brookfield
Plant Size	42.0 mgd	10.5 mgd	2.4 mgd	6.5 mgd	11.6 mgd	25.5 mgd	13.4 mgd	5.3 mgd	2.4 mgd
Plant Type	Step Aeration Activated Sludge	Trickling Filter	Trickling Filter	Trickling Filter	Activated Sludge	Activated Sludge	Activated Sludge	Activated Sludge	Contact Stabilization Activated Sludge
Phosphorus Removal Chemical(s)	$FeCl_3$, Polymer	$FeCl_3$, Polymer	$FeCl_3$, Polymer	$FeCl_3$, Polymer	$Al_2(SO_4)_3$, Polymer	$FeCl_3$, Polymer	$Al_2(SO_4)_3$	$Al_2(SO_4)_3$	$FeSO_4$
Point(s) of Chemical Addition	Tertiary	Secondary	Primary	Primary	Secondary	Primary	Secondary	Secondary	Secondary
Point of Combination of Chemical-Laden and Other Sludges	Anaerobic Digester	Gravity Thickener	Before Primary Clarifier	Before Primary Clarifier	Before Gravity Thickener	Before Primary Clarifier	Before Gravity Thickener	Sludges Not Combined	Before Primary Clarifier, In Aerobic Digester, or Before Chemical Conditioning
Sludge Treatment and Disposal Methods	Gravity Thickener, Anaerobic Digester, Air Drying on Land, Noncommercial Fertilizer	Gravity Thickener, Vacuum Filter, Fluidized Bed Incinerator	Anaerobic Digester, Drying Beds, Dump site	Thermal Conditioner, Vacuum Filter, Sanitary Landfill, Composting, Non-commercial Soil Conditioner	Gravity Thickener, Centrifuge, Fluidized Bed Incinerator, Ash Thickener, Ash Vacuum Filter	Anerobic Digester. Vacuum Filter, Incinerator	Anaerobic Digester, Flash Dryer, Drying Beds. Non-commercial Soil Conditioner, Liquid Sludge Hauling	Aerobic Digester, Dual Cell Gravity Concentrator, Agricultural Land Application	Pressure Filter, Multiple Hearth Incinerator

When the two upflow clarifiers were put into operation, one was started on lime and the other on ferric chloride for comparison. Tenech's predictions were borne out during the 18-mo qualification period before the U.S. Environmental Protection Agency approved the plant's performance and the city of South Bend's control. The plant immediately converted to full-scale operation with ferric chloride. By converting to ferric the savings in chemical costs amount to $720,000.00/yr. The following report will discuss the dramatic differences in sludge generation rates and sludge processing with lime and with ferric chloride.

South Bend is a well-managed plant with relatively few operational difficulties. It receives no abnormal industrial wastes because there is a minimal amount of industry now remaining in the community. Studebaker Motor Company quit its South Bend operations in the early 1960's. A brewery also closed down about 3 yr ago, reducing the plant influent BOD from approximately 150 to the present 70 to 80 mg/ℓ. Suspended solids were similarly reduced to about 80 mg/ℓ and digester gas production decreased somewhat.

There are no dramatic seasonal fluctuations in flow or waste loading. The major problem, because of combined sewers in the city, is tremendous hydraulic variation due to infiltration. The impact of high storm flows is that the plant influent is very dilute in BOD and SS, and it contains a great deal of grit and bitumen washed off the streets. During these periods, the TS content of the sludge pumped from the primary clarifiers is low, but subsequent thickening gets the solids concentration back to 4 or 5 percent before the sludge is pumped to the digester. A primary sludge degritter helps reduce the grit load on the digesters. Even so, it is necessary to clean one of the digesters each year on a rotating basis.

History

A description of plant modifications affecting sludge production and characteristics follows:

1956 - The original primary and secondary treatment plants were constructed to handle 90,840 m^3/day (24 mgd) average dry weather flow.

1974 - The facilities were modified and expanded to handle 181,680 m^3/day (48 mgd) average dry weather flow. The new equipment consisted of:

- An automated mechanical bar screen
- A primary sludge gravity thickener
- A sludge degritter
- Three waste-activated sludge centrifuges
- One additional aeration tank

- Two additional 6,586 m^3 (1.74 mil gal) secondary clarifiers
- A chemical sludge gravity thickener
- Three lime sludge centrifuges
- Two upflow clarifier tertiary treatment units.

April
1975 - Phosphorus removal was initiated with lime addition to one upflow clarifer and ferric and polymer addition to the other. The chemical sludges were combined, gravity thickened, centrifuged, and trucked to a storage area.

Sept.
1975 - The use of the waste-activated sludge centrifuges was discontinued. Waste-activated sludge was recirculated to the wet well ahead of primary clarifiers.

The use of the lime sludge centrifuges was discontinued. Chemical sludge was pumped to the sludge lagoons 4 km (2.5 mi) from the plant.

Oct.
1976 - Lime addition for phosphorus removal was discontinued. Ferric chloride and polymer were added to both of the upflow clarifiers. The chemical sludge was gravity-thickened and fed to the anaerobic digester.

Chemical Addition for Phosphorus Removal

Lime Addition--

Approximately 36.3 t/day (40 tons/day) lime as CaO were added to remove phosphorus in one of the upflow clarifiers. The average dosage was approximately 523 mg/ℓ.

The lime dosage was regulated by checking the pH of the water in the upflow clarifiers. The pH was maintained between 9.6 and 10.2. The ideal pH was considered to be 9.8. Above pH 10.2, the generation of sludge was dramatically increased due to the co-precipitation of magnesium hydroxide. Below pH 9.6, poorer phosphorus removal was achieved.

Ferric Chloride and Polymer Addition--

Liquid ferric chloride (37 percent $FeCl_3$) containing approximately 1.1 t/day (1.2 tons/day) $FeCl_3$, and 17 kg/day (38 lb/day) polymer are added to remove phosphorus in each clarifier. The average dosages are approximately 16 mg/ℓ $Fe Cl_3$ and 0.2 mg/ℓ polymer. The chemical feed rate is selected to achieve 85 percent reduction of the phosphorus concentration entering the plant. The feed rate is controlled by manual adjustment of the feed pump settings to compensate for changes in influent flow rate.

General Description of Wastewater Treatment Operations Affecting Sludge

Figure C-1 presents a general treatment plant flow diagram. Design parameters for the clarifiers and aeration basins are presented in Table C-2. Table C-3 presents a summary of 1976 influent flow characteristics and SS, BOD, and total phosphorus removals during primary, secondary, and tertiary treatments.

The raw wastewater is degritted by gravity in non-aerated grit chambers. Degritting reduces the grit load on the anaerobic digesters, increasing the volatile solids concentration of the digester feed and preventing grit accumulation in the sides and corners of the digesters.

Primary treatment removes only about 39 percent of the SS and 22 percent of the BOD entering the plant. The eight rectangular primary tanks have chain-type straight line collector arms for scraping the sludge into the end hoppers. The withdrawal of primary sludge from the hoppers is an automatic operation. Each of the eight clarifiers is pumped in series for three minutes in a continuous cycle.

Suspended solids and BOD removals during secondary treatment averaged 82 and 80 percent, respectively, in 1976. According to the plant chemist, this high degree of treatment is due to a low food-to-microorganism ratio in the mixed liquor. The activated sludge microorganisms are in a near-starvation phase so that they efficiently degrade the volatile material in the wastewater.

The five octagonal or circular secondary clarifiers have revolving nozzle-type chain belt collector arms which continuously scrape the sludge to the center hoppers. Sludge is continuously pumped from each hopper and either wasted to the wet well or returned to the aeration tanks. The amount of sludge which is returned averages 37 percent of the flow entering the plant.

Figure C-2 shows the configuration of the two 61-m (200-ft) diameter tertiary upflow clarifiers. Flow is directed from the secondary clarifiers via a 196-cm (77-in) line to a pair of gates which divert the flow into two parallel clarifiers. At this point, ferric chloride is dosed to serve as a flocculant. Flow entry to the clarifier is obtained through a 4.9-m (16-ft) diameter upward flow cone in the bottom of the center of each clarifier. Here a polymer is mixed into the upward flow and agitated by an eight-bladed propeller. The solids then gain entry into the flocculation zone which is surrounded by a shroud which serves as a baffle. The clarifier is designed so that the water recirculates within the shroud approximately eight times before it reaches the velocity at which it flows out from under it and down through the sludge blanket. As it flows through the sludge blanket, the flocculated solids are captured. The clarified water moves to the surface and exists over two concentric weirs.

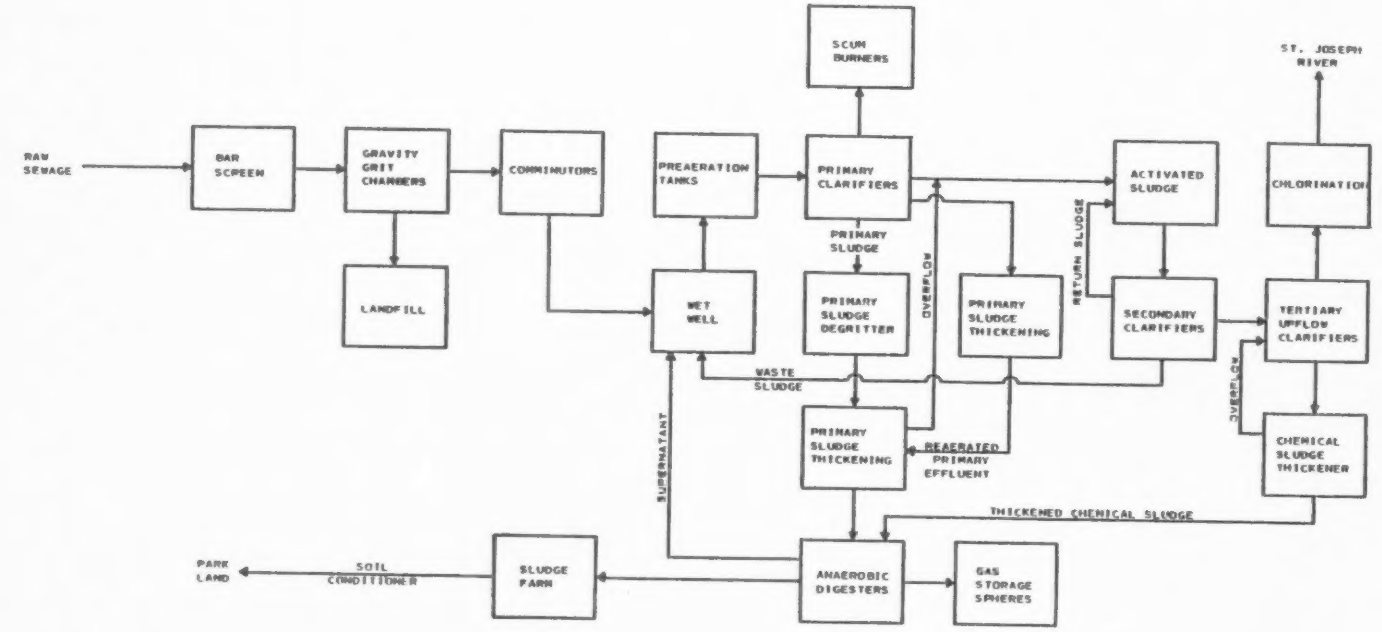

RAW SEWAGE
BAR SCREEN
GRAVITY GRIT CHAMBERS
LANDFILL
COMMINUTORS
PREAERATION TANKS
WET WELL
PRIMARY CLARIFIERS
SCUM BURNERS
PRIMARY SLUDGE
PRIMARY SLUDGE DEGRITTER
OVERFLOW
PRIMARY SLUDGE THICKENING
PRIMARY SLUDGE THICKENING
REAERATED PRIMARY EFFLUENT
WASTE SLUDGE
SUPERNATANT
ACTIVATED SLUDGE
RETURN SLUDGE
SECONDARY CLARIFIERS
TERTIARY UPFLOW CLARIFIERS
OVERFLOW
CHEMICAL SLUDGE THICKENER
THICKENED CHEMICAL SLUDGE
CHLORINATION
ST. JOSEPH RIVER
ANAEROBIC DIGESTERS
GAS STORAGE SPHERES
SLUDGE FARM
SOIL CONDITIONER
PARK LAND

Figure C-1. South Bend, Indiana, wastewater treatment plant flow diagram.

TABLE C-2. WASTEWATER TREATMENT PROCESS DESIGN PARAMETERS, SOUTH BEND, INDIANA

Unit	Description
Plant Design Flow Capacity	0.18 mil m^3/day (48 mgd) Avg. dry weather 0.30 mil m^3/day (80 mgd) Avg. wet weather
Primary Clarifiers (8)	Shape: Rectangular Size: 12.2 m x 36.6 m x 2.9 m SWD (40 ft x 120 ft x 9.5 ft SWD) Total Capacity: 10,214 m^3 (364,800 ft^3) Detention: 82 min @ Avg. dry weather flow
Aeration Tanks (4)	Type: Step Aeration Shape: Four Bay Rectangular Tanks Size: Each Bay 76.3 m x 7.3 m x 3.7 m SWD (250 ft x 24 ft x 12 ft SWD) Total Capacity: 32,260 m^3 (1,152,000 ft^3) Air Blower Capacity: 1,896 m^3/min (67,730 ft^3/min)
Secondary Clarifiers (5)	Shapes: 3 Octagonal, 2 Circular Sizes: 34.3 m x 3.7 m SWD (112.5 ft x 12 ft SWD) Octagonal Tanks; 44.2 m x 4.3 m SWD (145 ft x 14 ft SWD) Circular Tanks Total Capacity: 24,640 m^3 (6.51 mil gal) Detention: 192 min @ Avg. dry weather flow
Upflow Clarifiers (2)	Shape: Circular Size: 61.0 m x 4.9 m SWD (200 ft x 16 ft SWD) Total Surface Area: 5,840 m^2 (62,840 ft^2) Overflow Rate: 31.2 m^3/day/m^2 (766 gal/day/ft^2) @ Avg. dry weather flow Detention: 3.76 hr @ Avg. dry weather flow Total Capacity: 28,460 m^3 (7.52 mil gal)

TABLE C-3. SUMMARY OF 1976 WASTEWATER CHARACTERISTICS AND TREATMENT PERFORMANCE, SOUTH BEND, INDIANA

Population	
Connected	150,000
Equivalent BOD	156,643
Equivalent suspended solids	152,330
Wastewater Flow	
mil m^3/day (mgd)	0.17 (44.9)
mil m^3 treated	61.9 (16,378)
m^3 (gallons) per capita daily flow	1.13 (299)
Suspended Solids	
Influent, mg/ℓ	82
Primary effluent, mg/ℓ	49
Secondary effluent, mg/ℓ	9
Final effluent	8
% removal in primary	39
% removal in secondary	82
% removal in tertiary	17
% removal overall	90
BOD_5	
Influent, mg/ℓ	71
Primary effluent, mg/ℓ	56
Secondary effluent, mg/ℓ	10
Final effluent, mg/ℓ	2
% removal in primary	22
% removal in secondary	80
% removal in tertiary	79
% removal overall	97
Total Phosphorus	
Influent, mg/ℓ	2.29
Primary effluent, mg/ℓ	2.19
Secondary effluent, mg/ℓ	0.93
Secondary effluent, kg/day (lb/day)	158 (349)
Final effluent, mg/ℓ	0.24
Final effluent, kg/day (lb/day)	41 (90)
% removal in tertiary	74
% removal overall	88

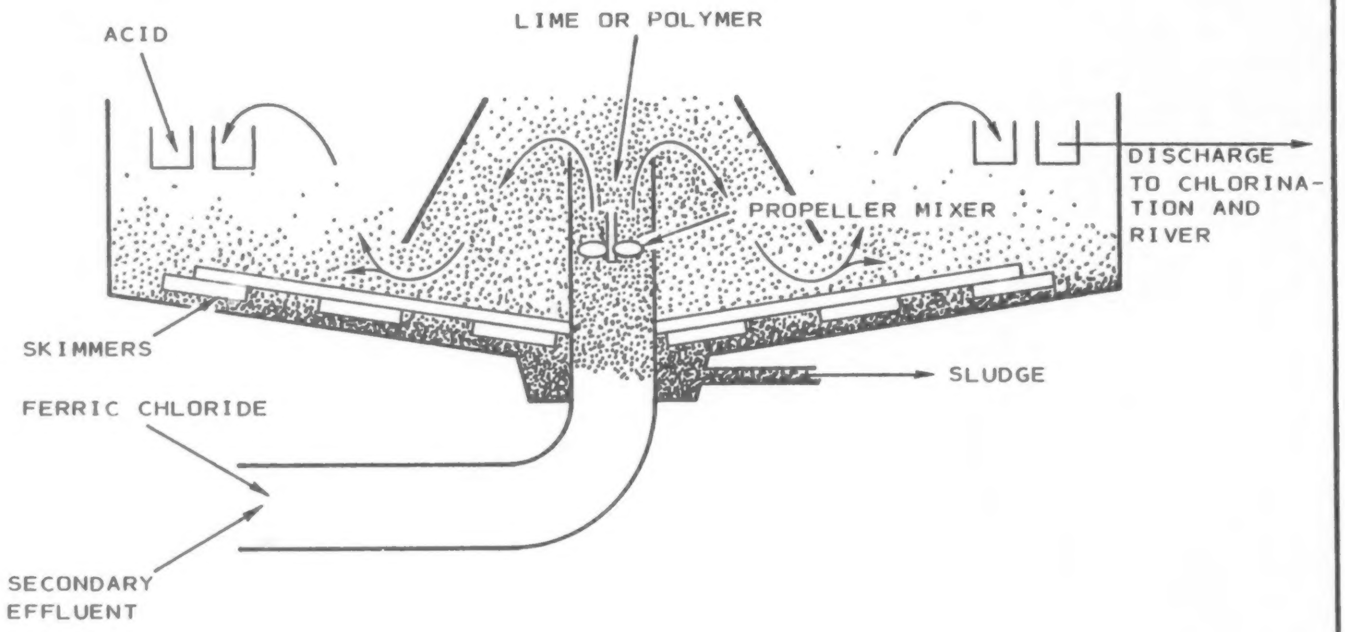

ACID
LIME OR POLYMER
PROPELLER MIXER
DISCHARGE TO CHLORINATION AND RIVER
SKIMMERS
SLUDGE
FERRIC CHLORIDE
SECONDARY EFFLUENT

Figure C-2. Tertiary upflow clarifier configuration, South Bend, Indiana.

When lime is used for phosphorus removal, it is added directly to the propeller mixing zone in the upflow cone of the clarifier. No polymer is needed. In contrast, polymer is necessary with ferric chloride to produce an ideally settling floc. The 244-cm (96-in) line leading to the center well acts as a long plug-flow reactor in which the ferric floc can develop to the correct size before it reaches the upflow cone. The low-velocity, large propeller mixer at that point can disrupt the floc or, with the addition of the polymer at that point, enhance the settling characteristics of the floc. The polymer is also necessary since without it temperature gradients produced by the sun passing over the clarifier during the day produce vertical water circulation and floc escapes. The polymer tends to weigh down the floc and prevent resulting instability of the sludge blanket.

The plant experiences some problems with the upflow clarifiers during periods of storm flow. To compensate for flow variations, the operator must regulate the speed of the variable-speed propeller mixer in the upflow cone. Speeding up the propeller tends to circulate the sludge solids in a tight circle with the shroud and prevents the velocity vectors from going outward.

Operational parameters for the upflow clarifiers during ferric and polymer addition have been determined by experience. An empirically determined relation between the SS concentration in the shroud and clarifier performance forms the basis for deciding when to pump sludge. The SS concentration is sampled once each day and maintained between 1,500 and 2,500 ppm. At greater than 2,500 ppm, sludge is pumped out of the center donut-like collection well. At less than 1,500 ppm, pumping is stopped and sludge allowed to accumulate. When the concentration goes as high as 4,000 ppm, the sludge blanket is too high and floc is lost over the weirs. When the concentration falls below 1,500 ppm, the sludge blanket is too thin and poor phosphorus removal results. Because of the low capacity of the sludge pumps, pumping is continuous from one clarifier or the other.

Clarifier operation with lime addition differs from operation with ferric in that sludge is continuously pumped from each clarifier. An efficient sludge pumping system is necessary for the heavy lime sludge -- sludge lines should have few elbows. In addition, when lime is used, reduction of the final effluent pH below 9.0 is necessary, requiring the addition of approximately 9.5 m^3 (2,500 gal)/day sulfuric acid to the effluent weir. On the other hand, because of the high pH produced with lime, no chlorine disinfection is necessary.

Detailed Description of Sludge Treatment and Disposal Operations

Table C-4 presents design parameters for the major sludge treatment steps at South Bend. The treatment and disposal of the lime/iron sludge produced at the plant in the past and the

TABLE C-4. SLUDGE TREATMENT PROCESS DESIGN PARAMETERS, SOUTH BEND, INDIANA

Unit	Description
Primary Sludge Gravity Thickener (1)	Shape: Circular Size: 18.3 m x 3.0 m SWD (60 ft x 10 ft SWD) Volume: 792 m^3 (28,280 ft^3) Surface Area: 263 m^2 (2,828 ft^2) Solids Loading: 28 kg/m^2/day (5.7 lb/ft^2/day)
Waste-Activated Sludge Centrifuges (3)	Type: Bird Continuous Scroll Solid Bowl Solids Feed Rate: 400 kg/hr (882 lb/hr) Hydraulic Feed Rate: 276 ℓ/min (73 gal/min) Centrate BOD_5: 3 to 8 mg/ℓ Centrate SS: 6.7 mg/ℓ
Anaerobic Digestion Tanks (4)	Type: 2-Primary, 2-Secondary Shape: 33.5 m x 7.8 m SWD + 2.7 m cone (110 ft x 25.5 ft SWD + 9 ft cone) Total Capacity: 30,280 m^3 (8 mil gal) Gas Storage Capacity: 7,000 m^3 @ 28,124 kg/m^2 (250,000 ft^3 @ 40 ps) Volatile Loading Rate: 0.6 kg VS/m^3/day (0.037 lb VS/ft^3/day)
Chemical Sludge Gravity Thickener (1)	Shape: Circular Size: 30.5 m x 3.0 m SWD (100 ft x 10 ft SWD) Surface Area: 730 m^2 (7,854 ft^2) Volume: 2,200 m^3 (78,540 ft^3) Solids Loading: 75.2 kg/m^2/day (15.4 lb/ft^2/day)
Lime Sludge Centrifuges (3)	Type: Bird Continuous Scroll Solid Bowl Hydraulic Feed Rate: 125 ℓ/min (33 gal/min)

iron sludge presently produced are described below. It should be recalled that the chemical sludges have always been gravity-thickened separately from the plant's organic sludge. Gravity-thickened sludge is presently anaerobically digested along with the organic sludges normally digested. The lime/iron sludge which was produced in the past, however, was never introduced into the digesters, but was centrifuged and trucked to a storage area.

A brief description of the procedures for handling the organic as well as the chemical sludges follows so that the digester performance can be related to sludge characteristics. As previously mentioned, waste-activated sludge is recirculated to the wet well ahead of the primary clarifiers. An estimated 55 percent of the solids removed from the primary clarifiers is derived from the recirculated secondary sludge. This percentage is rather high because of the high SS removal efficiency of the plant's secondary relative to its primary treatment operation.

Degritting--

Sludge from the primary clarifiers is pumped at less than one percent TS to allow effective degritting. Sludge degritting reduces the grit load on the anaerobic digesters, thus increasing the volatile solids concentration of the digester feed and preventing grit accumulation in the sides and corners of the digesters. The VS fraction of TS in the digester feed rose from approximately 62 percent to 65.5 percent of TS because of sludge degritting. A slight increase in digester gas production may have resulted.

Organic Sludge Thickening--

Degritted sludge at less than one percent TS is pumped to the gravity thickener. The gravity thickening step raises the TS concentration of the sludge before it is pumped to the digesters, so that digester space is not wasted and large amounts of digester supernatant are not formed and pumped back to the wet well. The average TS concentration of the thickened sludge which was pumped to the digesters was 4.7 percent in 1976. The VS fraction of TS averaged 67.5 percent. On the average, approximately 267 m^3/day (70,530 gal/day) thickened sludge were pumped from the thickener to the digesters in 1976. The sludge is continuously drawn off at 378 to 568 ℓ/min (100 to 150 gal/min).

Chemical Sludge Thickening--

The single chemical sludge gravity thickener is continuously fed with iron sludge at approximately 0.16 m^3/min (43 gal/min). When one of the tertiary clarifiers was operated with lime, sludge feed to the thickener was continuous at 0.89 m^3/min (234 gal/min). The thickened iron sludge is removed from the hopper at the bottom of the thickener once every 3 days. Pumping lasts for 7.8 hr at 0.38 m^3/min (100 gal/min). Formerly, the lime/iron sludge was pumped to the three lime centrifuges every day at a total rate of 0.48 m^3/day (126 gal/min) for 16 hr/day. A comparison of the amounts of raw and thickened sludge produced with iron and lime/iron is presented in Table C-5.

Table C-5 reveals that the mass of raw lime sludge produced by a single clarifier was 38 times the mass of iron sludge produced by both clarifiers. The TS concentration of the raw iron sludge is 1.0 percent, while that of the raw lime sludge was 4.0 percent. The combined lime/iron sludge was thickened to 10.2 percent TS. The amount of thickened iron sludge produced was 59 m^3/day (15,570 gal/day) by volume and 2.36 t/day (2.6 tons/day) by weight. In contrast, it is estimated that the amount of lime sludge which would have been produced if lime had been used in both tertiary clarifiers would be approximately 833 m^3/day (224,000 gal/day) by volume and 907 t/day (100 tons/day) by weight.

Anaerobic Digestion--

There are four 33.5-m (110-ft)-diameter anaerobic digestion tanks on-site with floating covers. The operation of the digestion tanks follows the pattern shown in Figure C-3.

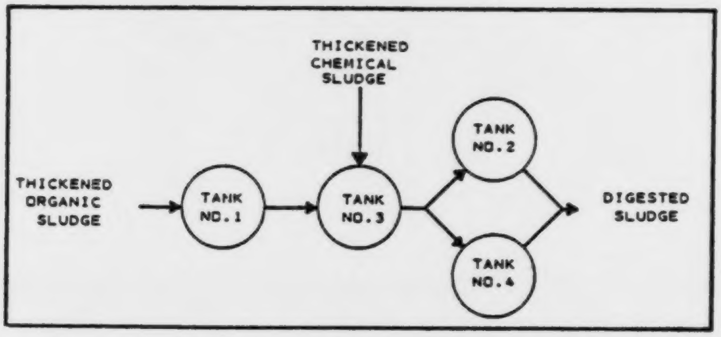


Figure C-3. Flow pattern for South Bend anaerobic digester.

Only the first two digestion tanks (No. 1 and No. 3) are heated and mixed. Mixing is accomplished with the Carter gas recirculation system. Each of the two digesters contains six Carter Aero-Hydraulic guns. Digester gas is pumped down vertical tubes to the bases of the guns, creating gas bubbles at the bottom which lift the sludge to the top of the tank and create the circulatory patterns in the tank. In the past, the plant had experienced severe upsets and souring of the digesters because of inadequate mixing. With the installation of the Carter System, operation has been excellent and there have been no upsets.

TABLE C-5. CHEMICAL SLUDGE PRODUCTION AND GRAVITY THICKENING, SOUTH BEND, INDIANA

	Iron Sludge	Lime Sludge	Lime/Iron Sludge
WASTEWATER TREATED			
mil m^3/day	0.14	0.07	0.14
(mgd)	(36.7)	(18.35)	(36.7)
RAW SLUDGE PRODUCTION			
m^3/day	236	1,129	N/A
(gal/day)	(62,280)	(298,200)	
% TS, dry weight	1.0	4.0	N/A
kg TS/mil m^3 wastewater treated	16,900	649,858	N/A
(lb/MG)	(141)	(5,420)	
THICKENED SLUDGE PRODUCTION			
m^3/day	59	N/A	454
(gal/day)	(15,570)		(120,000)
% TS, dry weight	4.0	N/A	10.2
VS, % of TS	35	N/A	11
kg TS/day	2,358	N/A	46,345
(lb/day)	(5,194)		(102,081)
kg VS/day	825	N/A	5,094
(lb/day)	(1,818)		(11,220)
t TS/day	2.36	N/A	46.26
(tons/day)	(2.6)		(51)

Note: N/A = Not Available

Digester heating is accomplished in two ways. Sludge from the No. 1 and No. 3 digesters is pumped through a heat exchange system. The boiler for the heat exchange system is run on methane digester gas. There is also a sludge preheating system of the heat-exchange type. Sludge from digester No. 1 is pumped to the preheating unit. The partially digested sludge mixes with the primary plus waste-activated sludge coming from the organic sludge gravity thickener. The digesting sludge feeds the raw sludge, and the mixture is heated. The preheater utilizes heat from cooling water coming off the plant's air compression engines, which are also run on methane digester gas.

Because of plant design, the thickened chemical sludge cannot also be preheated before being pumped into the digester. Therefore, it is at a lower temperature than the organic sludge and is capable of suppressing the temperature in the digester into which it is fed. This could result in a loss of digester gas production. Since most of the gas production takes place in the No. 1 digester, the chemical sludge is fed into the No. 3 digester. The result has been a 4° to 5° loss in temperature in this digester, but a decrease in total gas production has been avoided.

Intermittent pumping of organic sludge to the No. 1 digester occurs for 12 hr/day (2 out of every 4 hr), 7 days/wk. The primary digester receives an average of 267 m^3/day (70,530 gal/day). It receives 12,550 kg TS/day (27,650 lb TS/day) and 8,470 kg VS/day (18,660 lb VS/day). Pumping of iron sludge to the No. 3 digester occurs for 7.8 hr every third day. The average amount pumped is 59 m^3/day (15,570 gal/day). This contains approximately 2,358 kg (5,194 lb) of TS and 825 kg (1,818 lb) of VS. Some degradation of the VS contained in the iron sludge is occurring, as evidenced by the increase in digester gas production which has occurred. The level of digester gas production before and during addition of the iron sludge to the digester is shown in Table C-6. Iron sludge was first fed to the digester on 10/14/76. During iron sludge addition, digester gas production was higher by about 0.11 m^3 gas/kg VS destroyed (1.87 ft^3/lb VS). The level of digester gas production is still low compared to the nationwide average of 0.86 to 1.11 m^3 gas/kg VS destroyed (14 to 18 ft^3 gas/lb VS destroyed). It is possible that the waste-activated sludge has been so severely degraded as the result of the low food to microorganism ratio in the aeration basins that little volatile matter remains, causing the low gas production.

Some time ago, a fluoride dye tracer study of the digester was performed to calculate the actual liquid retention time based on an input of 378.5 m^3/day (100,000 gal/day). Instead of the expected 20 days, the dye tracer study indicated a shorter detention time of approximately 15 days.

TABLE C-6. ANAEROBIC DIGESTER GAS PRODUCTION SOUTH BEND, INDIANA

Month/ Year	1000 m^3 gas/day	1000 ft^3 gas/day	m^3 gas/kg VS destroyed	ft^3 gas/lb VS destroyed
1/75	5.26	188	0.54	8.7
2/75	5.12	183	0.45	7.3
3/75	5.12	183	0.49	7.9
4/75	5.10	182	0.44	7.1
5/75	5.66	202	0.53	8.6
6/75	4.79	171	0.52	8.5
7/75	4.00	143	0.47	7.7
8/75	3.70	132	0.42	6.9
9/75	4.68	167	0.57	9.3
10/75	5.18	185	0.60	9.8
11/75	5.29	189	0.59	9.6
12/75	4.73	169	0.49	7.9
1/76	4.48	160	0.55	9.0
2/76	5.54	198	0.58	9.4
3/76	4.84	173	0.57	9.2
4/76	5.18	185	0.60	9.8
5/76	4.56	163	0.57	9.3
6/76	5.15	184	0.59	9.5
7/76	4.62	165	0.56	9.1
8/76	4.23	151	0.50	8.2
9/76	5.26	188	0.59	9.6
10/76	5.40	193	0.69	11.2
11/76	5.40	193	0.67	10.9
12/76	5.71	204	0.73	11.9
1/77	5.38	192	0.73	11.8
2/77	5.80	207	0.67	10.9
3/77	5.40	193	0.61	9.9

Sludge is withdrawn from the No. 2 and 4 digestion tanks approximately every three days as space is needed in the No. 1 and 3 tanks. Sludge is removed from the No. 2 and 4 tanks through conventional bottom outlets. Withdrawal from the No. 1 and 3 tanks is through "spider drawoffs," three pipes near the bottom which extend toward the center of the tank and branch at the ends into three arms. The spider which is used to draw off sludge is rotated each time.

The digested sludge removed from the digester averages 5.5 percent TS with a VS fraction of 47.2 percent. Iron sludge addition had no effect on the TS and VS concentrations of the digested sludge. The average volume of sludge removed from the digester was 118 m^3/day (31,300 gal/day) before iron sludge addition and 160 m^3/day (42,200 gal/day) during iron sludge addition. So the volume of digested sludge increased by 42 m^3/day (10,900 gal/day). During the same period, the volume of supernatant increased by 8 m^3/day (2,100 gal/day).

Disposal of Digested Sludge--

Digested sludge is pumped through a 15-cm (6-in) steel pipeline from the digester directly to a "sludge farm" which is 4 km (25 mi) away. The sludge is pumped with a 100-HP, 126-ft pumping head, centrifugal pump. At the farm, the sludge can be routed to any one of seven different fields of varying size. The diked fields are filled with approximately 76 cm (30 in) of sludge. Approximately 757 m^3 (200,000 gal) of sludge are pumped to a field at any one time. It takes approximately 2 to 3 mo to fill a field. The sludge running at 5 to 6 percent TS is allowed to dewater on the fields through percolation and evaporation for a time before a No. 4230 John Deere tractor of special design mixes up the sludge and exposes wet section to the air. There are only a few operators who are capable of driving this tractor, since the front end has a capacity to lift up, depending on the field conditions. A weight is added to the front end for better balance. The front wheel spread can also be varied to achieve any width desired. The wheels on the rear are specially designed cleat-type steel wheels capable of tracking through sludge and mud.

The 91 ac at the site were purchased around 1967. The site became operational that same year. The fields are effectively screened from the roadway by a tract of trees approximately 0.2 mi in width. The seven fields are located on the back of the property.

Sludge is pumped to the fields year round, but normally worked with the tractor only in the summer. Depending on weather conditions and other variables, the sludge fields are worked with the John Deere tractor to produce a relatively dry sludge cake. The tractor wheels work down as far as 30 to 38 cm (12 to 15 in) into the soil and sludge to work it into a homogenous mixture. After 18 mo, the material is removed with a front-end payloader and stockpiled elsewhere on the property. Last year, for example, contractors removed almost all of the stockpiled sludge and used it on various construction projects around South Bend.

Four lagoons were originally constructed on the site for receipt of digester wash water. Treated effluent is used in cleaning out the digesters. This wash water contains grit, detritus and residual sludge having some BOD. Odors from the lagoons do not seem to be a problem. The No. 4 lagoon is now being filled in with rubbish and covered over. Some of the dried soil-sludge mixture produced at the site will be used as surface cover on this lagoon.

The site supposedly has a clay layer underneath it. However, in an exposed section being dug out for fill material and for making a deeper field for later sludge filling, there were 0.6 to 0.9 m (2 to 3 ft) of surface soils underlain by approximately 3.0 to 3.7 m (10 to 12 ft) or more of silty sand. The spreading area is part of a large meandering ancient stream or flood plain area.

Lime/Iron Sludge Dewatering and Disposal--

When the plant was using lime to remove phosphorus from half of the wastewater, the thickened lime/iron sludge produced was dewatered by centrifuges. The thickened sludge, at 11 percent TS, was fed to each of the three centrifuges at a rate of 159 ℓ/min (42 gal/min). Because of the large volume of lime/iron sludge to be dewatered, it was necessary to run all three centrigues for approximately 16 hr/day, 7 days/wk. If the plant had used lime to remove phosphorus from all of the wastewater, the volume of lime sludge would have been almost doubled, and this amount could not have been centrifuged.

There were several problems with operation of the centrifuges. The first problem has been mentioned: the total centrifuge capacity was inadequate to handle all of the sludge. The design engineers estimated that only 27 t/day (30 tons/day) of dry sludge TS would be produced by the two upflow clarifiers operating with lime, while the actual production figure for both clarifiers is closer to 91 t (100 tons). Other operational problems resulted because of the poor instructions given to the plant on centrifuge operation and maintenance procedures. The heavy chemical sludge caused breaking of shear pins, and frequent equipment overhauls were necessary. Maintenance costs were high during the 6-1/2 mo that the centrifuges were in operation.

The lime/iron sludge was centrifuged to a cake that was 43 percent TS, and 8 percent of the solids were volatile. The cake was trucked out to a storage site. This disposal method was not satisfactory and was only meant to be temporary until the plant stopped using lime for phosphorus removal. One of the problems with this method was the spillage of lime from the truck onto the road, discoloring the pavement.

Sludge Treatment and Disposal Costs

The best cost data available from plant record is related to anaerobic digestion. Unfortunately, the operational costs for chemical sludge thickening and centrifugation cannot be broken apart from the operational costs of handling phosphorus removal chemicals and tertiary upflow clarifier operation. Because no special problems were involved in the chemical sludge thickener operation, the costs can be assumed to be similar to those of a similar organic sludge thickening operation. Of course, the total cost to thicken per ton of dry solids is related to the mass of sludge treated. For instance, the total cost of lime sludge treatment at South Bend would be much less per ton of dry solids than the total cost of treating iron sludge. This results from the mass of lime sludge being much greater than the mass of iron sludge. The size of the thickener, however, remains the same in each case, and it is run for the same number of hours each day.

The disadvantage of treating lime sludge comes in the disposal stage. At South Bend, it would not have made sense to send the lime sludge to the digesters and the sludge farm, as can be done with the iron sludge. While there is probably adequate room in the four digestion tanks to handle the large amount of lime sludge, the digesters would likely serve mainly as holding tanks for this sludge because of the inability to preheat the sludge and the low volatile content.

Pilot plant studies have shown that the lime sludge can be compacted in the thickener to as high as 16 percent TS, and even at this high concentration, the sludge can be pumped fairly easily.

It would be possible, then, to pump the lime sludge to the sludge farm, but ultimate disposal of the lime sludge would still be a problem because of its volume. Parkland disposal would not use up all of the sludge. Direct trucking of thickened and perhaps centrifuged lime sludge to agricultural land seems to be the best disposal alternative. Dewatering the sludge to a high solids concentration would lower the trucking costs. Pilot tests showed that the lime sludge centrifuges can operate with a feed sludge solids concentration of up to 30 percent TS. By feeding sludge at 16 percent TS, a fairly dry cake could be produced which would lower trucking costs.

The following costs related to digester operation were available from plant records.

Item	1975	1976
Operational Labor (4 full-time operators)	$74,981.18	$71,012.53
Maintenance of Equipment and Structures	$33,458.26	$35,235.74
Electricity	$10,462.40	Not Available
Miscellaneous	$10,903.11	$4,164.04

The capital cost for the four digestion tanks was $5,059,200 in 1956. This does not include the cost of the Carter sludge mixing system which was added later. The cost of the two sludge heat exchangers for tanks No. 1 and 3 plus piping and valve controls, and the sludge pump which pumps sludge from the No. 1 digester to the preheater, was $63,000 in 1956. The two digester gas storage spheres cost $225,000 each.

The acquisition costs for 91 ac of farmland were $100,800 in 1969. Land is very available near the treatment plant. The capital cost expenditures related to farm operations included purchase of the land; site development costs, including road improvements; installation of piping for the seven fields, including cross-connections, valves, boxes, and miscellaneous tubing and hoses; construction of a 4-km (2.5-mi), 15.24-cm (6-in) steel pipeline; and purchase of one John Deere Model 4230 special tractor, one front-end payloader, two multiple harrow plows, one rototiller disc harrow, and two special pieces of drag equipment. The operating costs of farm operation disposal have been estimated at $1.60/t ($1.45/ton) of dry sludge TS.

The additional costs for sludge treatment resulting from iron addition at South Bend were very minor, since there was plenty of room available in the digester. The chemical sludge increased the volume of digester feed by 22 percent and the mass of feed TS by 19 percent. The capital and operating costs of the chemical sludge gravity thickener, including pumping to and from, are the major additional costs.

Summary and Conclusions

A comparison of tertiary upflow clarifier operation with lime and with ferric chloride showed that the mass of raw lime sludge generated per mil m^3 of wastewater treated was 77 times that of the raw iron sludge produced. The TS concentration of the raw lime sludge was 4.0 percent. Gravity thickening of the lime sludge mixed with a relatively insignificant amount of iron sludge produced a TS concentration of 10.2. The sludge could be centrifuged to 43 percent TS. The centrifuged sludge cake was trucked to a storage area until difficulties arose regarding complaints along the route travelled by the trucks. Then the thickened sludge was pumped through a 4-km (2.5-mi), 15.24-cm (6-in) pipeline to lagoons located at the plant's secluded sludge farm. This was the mode of disposal until the plant discontinued using lime. The plant feels that the most satisfactory method of lime sludge disposal would be land application of thickened or centrifuged cake. Although there were problems with centrifugation - high maintenance and electricity costs for centrifuge operation and insufficient capacity provided by the design engineers - the centrifuges operated with feed concentrations of up to 30 percent TS and produced a correspondingly dry cake in pilot tests. These tests also showed that the lime sludge could be compacted in the gravity thickener to 16 percent TS and even at that high concentration it was easily pumped and flowed well.

The thickened iron sludge was fed to the anaerobic digester. By feeding it to the second of the two primary digesters which follow in series, a suppression of total gas production was avoided. This suppression could have occurred because of the plant's inability to preheat the chemical sludge as it does its organic sludge. Some temperature suppression occurred in the second digester as the result of chemical sludge feed, but most of the gas is produced in the primary digester, so the impact was not great. Other than the fact that a total decrease did not occur, the effect of the iron sludge on gas production is uncertain. Initially, there appeared to be an increase in the amount of gas produced per quantity of VS destroyed, but later this trend disappeared.

Digested sludge was pumped out to a sludge farm where it was dried and mixed with topsoil in fields with a special tractor and then given away to contractors and the city of South Bend for parks. The sludge farm appears to be an excellent concept for areas where land is available. The concept was developed by South Bend's Manager of Sanitary Operations who is a former farmer. The operating cost of the sludge farm operation is estimated at $1.60/t ($1.45/ton) of dry sludge TS.

CASE STUDY D: SHEBOYGAN, WISCONSIN

Introduction

Sheboygan provides an example of secondary addition of ferric chloride. It also provides an example of vacuum filtration of undigested (raw) ferric sludge. The data from this plant can be compared with data from plants, such as Midland, Michigan, and Windsor, Ontario, which have practiced vacuum filtration of raw ferric sludges produced by primary addition of the chemical.

Sheboygan operates a fluidized bed incinerator. The impact of ferric addition on energy requirements, maintenance, and repair requirements for this type of system can be examined at the plant. The performance can be compared with that of other plants such as in Brookfield, Wisconsin, where a ferric sludge undergoes pressure filtration and multiple hearth incineration.

Overall, Sheboygan is representative of the many older wastewater treatment plants which are awaiting future expansions and are overloaded both hydraulically and with respect to sludge handling capacity in the meantime. Because of the hydraulic loading problems, the plant is fortunate that only 4 percent of the sewers are storm rather than sanitary sewers. Industrial waste discharges contribute 40 percent of the flow to the plant. Included are industrial phenols, plastics, and leather tanning waste. Whether because of hydraulic conditions or because of industrial wastes, SS and BOD removal are poor. Table C-7 below presents some historical data on plant influent characteristics and suspended solids and BOD removal. The plant's nominal design treatment capacity is 0.060 mil m^3/day (16 mgd) average daily flow with 0.83 mil m^3/day (22 mgd) maximum flow. Its discharge permit allows effluent concentrations of 80 mg/ℓ SS and 70 mg/ℓ BOD. Phosphorus removal performance is also poor due to both hydraulic conditions and the lack of sludge handling capacity.

TABLE C-7. A SAMPLE OF HISTORICAL DATA INDICATING AVERAGE PLANT INFLUENT CHARACTERISTICS, SUSPENDED SOLIDS AND BOD REMOVAL, SHEBOYGAN, WISCONSIN

	April-Sept 1970	April-Sept 1973	April, May, July, Sept, Oct, Nov 1976
Influent m^3/day (mgd)	45,800 (12.1)	54,900 (14.5)	14,300 (10.9)
Influent SS (mg/ℓ)	202	184	195
Effluent SS (mg/ℓ)	34	53	48
Influent BOD (mg/ℓ)	196	211	219
Effluent BOD (mg/ℓ)	58	81	58

History

The following describes historical plant modifications affecting sludge production and characteristics.

1968 - Began sludge treatment with Dorr-Oliver vacuum filtration and Fluo Solids incineration system. Old anaerobic digesters converted to holding tanks for excess liquid sludge. Excess sludge applied to drying beds.

Jan. 1972 - Began phosphorus removal by chemical addition.

Feb.-Apr. 1972 - Discontinued use of incinerator heat exchanger. Excess liquid sludge hauled to farmlands instead of applied to drying beds.

Apr. 1973 - Began polymer addition to thickener.

Jun. 1975 - Thickener overflow recirculated to head of the primary tanks instead of before the trickling filters.

Oct. 1975 - Began keeping records of phosphorus removal chemical dosages and influent and effluent phosphorus concentrations.

Jan. 1977 - Began proportional addition of polymer to thickener feed lines.

Chemical Addition for Phosphorus Removal

In order to remove phosphorus at this treatment plant, liquid ferric chloride is added to a division well which distributes the effluent from the trickling filters to the two final clarifiers. Anionic polymer (Hercules 831) is then added to the center feed wells of the final clarifiers. The average ferric chloride dosage was 30 mg/ℓ dry $FeCl_3$ in 1976, with polymer dosage average 208 kg/mil m^3 (1.75 lb/MG). Because accurate records were not kept before 1976, chemical dosages before that time are unknown.

The reduction of the total phosphorus concentration in the wastewater averaged only about 54 percent in 1976. Average influent and effluent concentrations were approximately 7.02 ppm and 3.26 ppm total phosphorus, respectively. Phosphorus removal was very poor on the average, partly because during this period ferric chloride addition was stopped whenever thickener performance deteriorated. This occurred frequently. In 1977, when thickener performance was improved and ferric chloride addition rarely had to be stopped, average phosphorus removal increased to 66.5 percent. Influent and effluent phosphorus concentrations averaged 10 ppm and 3.3 ppm, respectively. The plant is required to reduce its effluent phosphorus concentration below 4 ppm to meet its discharge permit requirements. Eighty percent removal cannot be achieved at the plant because of its design. In order to get better removal, more contact time is needed between the addition of ferric chloride and polymer, but this cannot be achieved with the present plant design.

General Description of Wastewater Treatment Operations Affecting Sludge

Figure C-4 presents a general treatment plant flow diagram. The plant has no facilities for grit removal from the raw sewage. Primary settling occurs in two primary clarifiers having a total capacity of 5,680 m^3 (1.5 MG). Thickener overflow, vacuum filter filtrate, and incinerator scrubber water join the raw sewage before it enters the primary clarifiers. Occasionally, when the thickener is severely overloaded and cannot accept it, secondary sludge is pumped to the primary clarifiers. Primary sludge is continually removed from the bottoms of both primary clarifiers and pumped to the thickener by way of the primary sludge degritter.

Primary effluent is pumped over two rock-filled trickling filters. The trickling filter effluent is split and enters two 10.7-m (35-ft) diameter final clarifiers. Secondary sludge is removed continually from the bottoms of both clarifiers and pumped directly to the thickener.

Detailed Description of Sludge Treatment and Disposal Operations

Degritting--

Because there is no grit removal from the raw sewage, primary sludge degritting is essential. Grit lowers the volatile content of the sludge fed to the incinerator and increases the amount of fuel required to burn it. The Dorr-Oliver cyclone sludge degritter efficiently degrits primary sludge of less than one percent solids concentration.

Thickening--

Degritted primary sludge and sludge from the final clarifiers enter a single Dorr-Oliver gravity thickener through separate 15.2-cm (6-in) feed lines. The thickener is 12.2 m (40 ft) in diameter, and its depth is 3.05-m (10 ft) liquid sidewal plus the cone at the bottom. It is equipped with scraper arms having vertical pickets for continuously stirring sludge in the thickener as it is scraped over the bottom toward the center. Sludge is sucked into a sludge pape through eight portholes in the base of the thickener center column. It is then pumped to the vacuum filters when they are operating. On weekends, when the filters not operating, the operator has the option of pumping sludge to the holding tanks for later land application or not pumping at all.

Figure C-5 presents combined hydraulic and mass balances constructed around the thickener. It presents information on average sludge flow rates, total and volatile solids concentrations, and masses obtained from monthly reports. It is based on monthly averages of data for a 6-mo period of operation with no ferric addition and a similar later period with ferric addition.

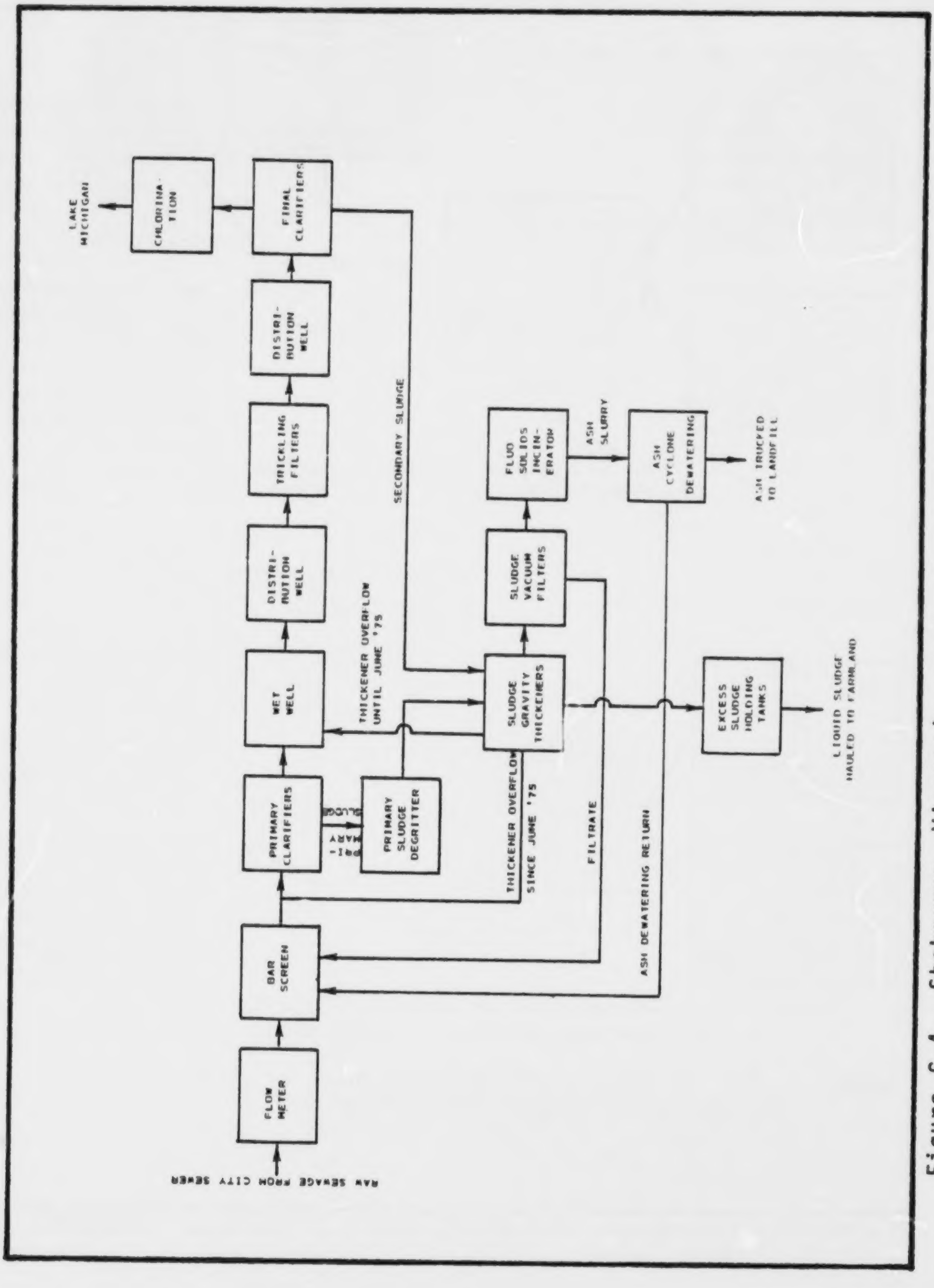

RAW SEWAGE FROM CITY SEWER
FLOW METER
BAR SCREEN
PRIMARY CLARIFIERS
PRI-MARY SLUDGE
PRIMARY SLUDGE DEGRITTER
WET WELL
DISTRI-BUTION WELL
TRICKLING FILTERS
DISTRI-BUTION WELL
FINAL CLARIFIERS
CHLORINA-TION
LAKE MICHIGAN
SECONDARY SLUDGE
THICKENER OVERFLOW UNTIL JUNE '75
THICKENER OVERFLOW SINCE JUNE '75
SLUDGE GRAVITY THICKENERS
SLUDGE VACUUM FILTERS
FLUID SOLIDS INCIN-ERATOR
ASH SLURRY
ASH CYCLONE DEWATERING
ASH TRUCKED TO LANDFILL
FILTRATE
ASH DEWATERING RETURN
EXCESS SLUDGE HOLDING TANKS
LIQUID SLUDGE HAULED TO FARMLAND

Figure C-4. Sheboygan, Wisconsin, wastewater treatment plant flow diagram.

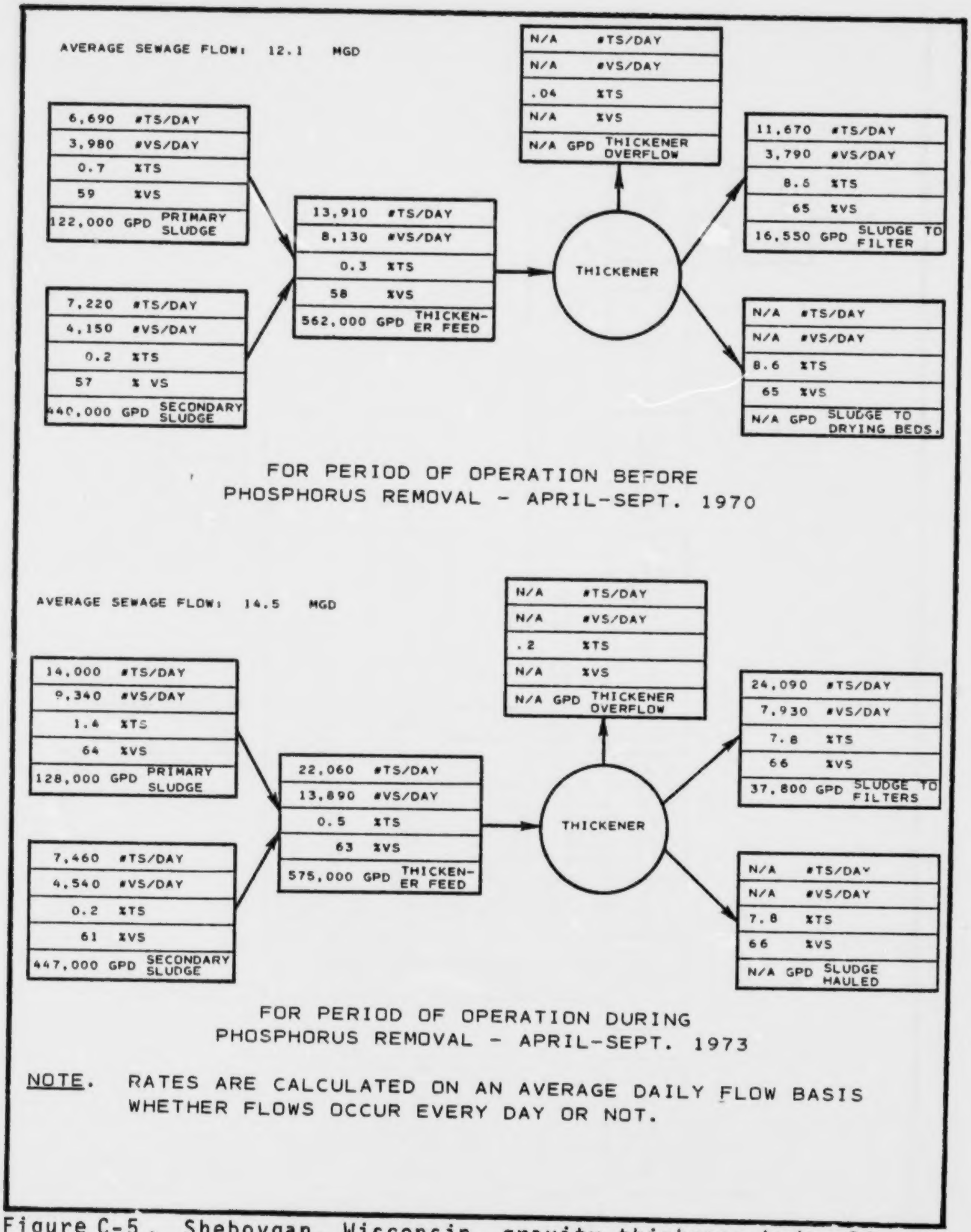

AVERAGE SEWAGE FLOW: 12.1 MGD
6,690 #TS/DAY
3,980 #VS/DAY
0.7 %TS
59 %VS
122,000 GPD PRIMARY SLUDGE
7,220 #TS/DAY
4,150 #VS/DAY
0.2 %TS
57 % VS
440,000 GPD SECONDARY SLUDGE
13,910 #TS/DAY
8,130 #VS/DAY
0.3 %TS
58 %VS
562,000 GPD THICKENER FEED
THICKENER
N/A #TS/DAY
N/A #VS/DAY
.04 %TS
N/A %VS
N/A GPD THICKENER OVERFLOW
11,670 #TS/DAY
3,790 #VS/DAY
8.6 %TS
65 %VS
16,550 GPD SLUDGE TO FILTER
N/A #TS/DAY
N/A #VS/DAY
8.6 %TS
65 %VS
N/A GPD SLUDGE TO DRYING BEDS.
FOR PERIOD OF OPERATION BEFORE PHOSPHORUS REMOVAL - APRIL-SEPT. 1970
AVERAGE SEWAGE FLOW: 14.5 MGD
14,000 #TS/DAY
9,340 #VS/DAY
1.4 %TS
64 %VS
128,000 GPD PRIMARY SLUDGE
7,460 #TS/DAY
4,540 #VS/DAY
0.2 %TS
61 %VS
447,000 GPD SECONDARY SLUDGE
22,060 #TS/DAY
13,890 #VS/DAY
0.5 %TS
63 %VS
575,000 GPD THICKENER FEED
THICKENER
N/A #TS/DAY
N/A #VS/DAY
.2 %TS
N/A %VS
N/A GPD THICKENER OVERFLOW
24,090 #TS/DAY
7,930 #VS/DAY
7.8 %TS
66 %VS
37,800 GPD SLUDGE TO FILTERS
N/A #TS/DAY
N/A #VS/DAY
7.8 %TS
66 %VS
N/A GPD SLUDGE HAULED
FOR PERIOD OF OPERATION DURING PHOSPHORUS REMOVAL - APRIL-SEPT. 1973
NOTE. RATES ARE CALCULATED ON AN AVERAGE DAILY FLOW BASIS WHETHER FLOWS OCCUR EVERY DAY OR NOT.

Figure C-5. Sheboygan, Wisconsin, gravity thickener hydraulic and mass balance.

The hydraulic-mass balances show that during both periods, thickener performance, as measured by the solids concentration of the thickened sludge, was good. During the first period, part of 1970, with no ferric addition, a combined primary and secondary sludge of about 0.3 percent solids was thickened on the average to 8.6 percent TS. During the 1973 period, with ferric addition, the sludge feed averaged 0.46 percent TS and was thickened on the average to 7.76 percent TS. The thickened sludge solids concentration was lower in 1973, despite a higher concentration in the feed sludge and despite the addition of polymer to the sludge feed lines. An anionic polymer was used beginning in April of 1973, to aid sludge settling and to maintain a clear overflow. Problems with floating of the sludge blanket and resulting high overflow solids concentration were experienced in thickener operation both prior to and after phosphorus removal. They seem related to industrial waste discharges. Poor settleability was more frequent during ferric addition than before. In 1970, the average overflow solids concentration was 441 mg/ℓ TS, while in 1973, during ferric addition, it was 1,764 mg/ℓ. The plant operators responded to thickener upsets by stopping ferric addition and discontinuing sludge feed whenever thickener upsets occurred. Secondary sludge was wasted to the primary tanks at these times and the bottoms of these tanks were used for storage until the thickener regained stability. As a result, the primary sludge solids concentration and sludge mass was greater in 1973.

The total mass of primary and secondary sludge pumped to the thickeners was higher in 1973 by about 44,400 kg/mil m^3 (370 lb TS/MG). This amounts to an increase of approximately 2.18 t (2.4 tons) of dry sludge each day at a 49,200 m^3/day (13 mgd) influent flow rate. The total volume of sludge pumped to the thickeners was actually lower in 1973, however, because of the higher solids concentration.

The thickener has an overflow rate of approximately 18,300 to 18,700 ℓ/day/m^2 (450 to 460 gal/day/ft^2). A sludge blanket of at least 1.22 m (4 ft) must be maintained in the thickener to avoid pumping a thin sludge. The blanket depth is normally maintained at close to 1.52 m (5 ft).

The anionic polymer which is used in the thickener is the same one used for phosphorus removal (Hercules 831). Since polymer addition began, in April 1973, many different points of addition and dosages have been tried. It was added, for instance, to the secondary sludge feed line about 12.2 m (40 ft) before it enters the thickener, providing a 2- to 3-min contact time. Addition directly to the center well through a plastic hose was also tried both with and without a diffuser on the end of the line to distribute the polymer. None of the methods tried produced a stable sludge blanket. In January of 1977, a successful method was found. The method involves adding polymer separately to the primary and secondary sludge feed lines to achieve the same concentration of polymer in the sludge in each line. In other words, the amount of polymer added to each line is proportional to the sludge flow through the line. The polymer is added to each feed line through a plastic hose. The point of addition is immediately before the feed lines discharge in the center feed well of the thickener. The usual dosage is 3 ppm.

Since January of 1977, when proportional polymer addition was started, the thickener overflow has had a very low average TS concentration. The average for January through April was only 154 mg/ℓ.

Sludge is pumped from the thickener to either the holding tanks or the vacuum filters. The amount of sludge pumped to the holding tanks has not been recorded until recently. During six months from September 1976 through February 1977, sludge was pumped to the holding tanks for an average of 28.4 hr/wk. It is pumped during the 33 hr on weekends when the vacuum filters are not operating. The average number of gallons pumped per day was 7,535 and the mass of solids per day pumped was 2,000 kg (4,400 lb). During the same period, the average volume pumped to the vacuum filters was about 127,000 ℓ/day (34,000 gal/day) and the sludge mass filtered was 8,810 kg/day (19,400 lb/day).

Unfortunately, the hydraulic-mass balances in Figure 2 do not show how phosphorus removal affected the quantity of sludge pumped to the vacuum filters, or the amount of excess pumped to the holding tanks. The data necessary to calculate these quantities are unavailable. The increased amount of sludge which Figure C-5 shows was pumped to the filters in 1973 and was a result of longer operation of the filters and incinerator. This longer operation was a result of the elimination of certain problems experienced when the incinerator was new. Since 1972, the incinerator has been operated on its present schedule.

Liquid Sludge Hauling--

Liquid sludge from the holding tanks is trucked out and applied to agricultural lands by a private contractor. The contractor charges by the gallon for this service. The number of gallons hauled is not the same as the number of gallons pumped to the holding tanks. The tanks are not mixed, so the sludge becomes heavy in them. It must be diluted with water during pumping so that it will flow into the truck. Data on the quantities of sludge hauled in 1970 and 1973 are unavailable. In 1976, 22,700 m^3 (6 MG) were hauled, but only approximately 10,400 m^3 (2.75 MG) were pumped to the holding tanks.

Dewatering--

Sludge dewatering is accomplished with two vacuum filters which were installed in 1957 and rebuilt in 1968 with Dorr-Oliver components to be compatible with the Dorr-Oliver Fluo Solids incineration system. The filters have a design filtering capability of 454 kg (1,000 lb) TS/hr each at a feed solids concentration of 27 percent TS. The two rotary drum filters each have a filtering area of 18.6 m^2 (200 ft^2), face width of 2.44 m (8 ft), and diameter of 2.44 m (8 ft). The filter medium is polyethylene cloth. The filter drum speed is 7 min/revolution. This is the slowest drum speed setting possible. Sludge is conditioned prior to filtration with Hercules 814 cationic polymer and ferric chloride. The vacuum filters are operated for 24 hr/day on Monday through Friday, 15 hr/day on Saturday, and are shut down on Sunday.

Table C-8 presents a comparison of filter performance before and after phosphorus removal. Data on incineration rate and incinerator fuel consumption are also presented. The data are based on monthly reports for the 1970 "before phosphorus removal" and 1973 "after phosphorus removal" time periods previously mentioned. In addition, data from a similar 6-mo period in 1976 is presented. The plant manager felt that the 1976 data were more representative of the results after phosphorus removal than the 1973 data.

TABLE C-8. VACUUM FILTRATION AND INCINERATION PERFORMANCE BEFORE AND AFTER PHOSPHORUS REMOVAL, SHEBOYGAN, WISCONSIN

	1970	1973	1976
Filter yield, kg TS/m^2/hr (lb TS/ft^2/hr)	20.3 (4.16)	16.1 (3.3)	12.8 (2.62)
Filter feed solids % TS	8.6	7.76	7.0
Cake dryness % TS	25.5	24.7	21.5
Cake volatile VS (% of TS)	65.45	65.9	65.4
Filtrate solids ppm SS	658	383	442
814 Polymer, kg/t (lb/ton)	1.91 (3.82)	1.28 (2.55)	1.20 (2.4)
$FeCl_3$, kg/t (lb/ton)	NA	NA	66.3 (132.5)
Incinerator feed rate, kg/hr (lb) hr*	755 (1,662)	578 (1318)	475 (1,046)
Fuel consumption, l/t (gal/ton)	246 (59)	451 (108)	517 (124)

* Dry solids fed to incinerator based on solids in filter feed. Amount of solids lost in filtrate unknown.

Before phosphorus removal, ferric chloride was used for conditioning less than 20 percent of the time. Most of the time, polymer was used alone. After phosphorus removal, cake dryness was usually only about 18 percent TS without ferric conditioning; but with ferric conditioning, a 21-percent TS cake could be achieved.

Though ferric conditioning raised cake solids content, it did not improve filter yield. Filter yield can be raised by increasing filter drum speed, but at the expense of cake dryness.

The volatile content of the sludge cake does not appear to have decreased with phosphorus removal. Filtrate quality also was not adversely affected. Polymer dosage decreased somewhat as ferric chloride usage became regular.

Incineration--

The plant operates a Dorr-Oliver "F/S System" Fluo Solids incinerator. It has a design capacity of 122 kg/hr (2,684 lb/hr) dry filter cake. The cake is expected to contain 25 percent TS with 73 to 74 percent volatile content. The incinerator's overall height is 12.2 m (40 ft) and the cylinder's inside diameter is 3.05 m (10 ft). The expanded sand bed is 1.83 to 2.13 m (6 to 7 ft) deep and contains 7.26 t (8 tons) of silica sand. A construction plate supports the bed and allows fluidizing air to pass upward. It is a self-supported refractory arch with metal tuyeres which pass the air. The windbox chamber is below this. The windbox size is 1.07 m by 2.74 m (3.5 ft by 9 ft). A preheat burner mounted on the windbox preheats the fluidized bed in the reactor prior to sludge feed. A preheat burner fuel oil pump furnishes No. 2 fuel oil to preheat the burner. An atomizing blower supplies pressurized air to atomize the fuel oil in the preheat burner. Once the reactor bed temperature is preheated to 621°C (1,150°F), which is the auto-ignition temperature of sludge, No. 2 fuel is injected directly into the bed by the bed guns. The preheat burner is turned down but continues to serve as the method of heating the fluidizing air which is discharged through the windbox into the bed. The fluidized air comes from a centrifugal blower which discharges air at 4.5 to 5.0 psig into the reactor. Before April 21, 1972, the fluidizing air was heated by a heat exchanger rather than by the preheat burner. Problems with the heat exchanger prompted its discard, and the preheat blower became the method of fluidizing air heating. Without the heat exchanger, oil consumption is higher by about 129 ℓ/t (31 gal/ton). Without the heat exchanger, the incinerator capacity is reduced by about 159 kg/hr (350 dry lb/hr). If the sludge feed rate is increased above its capacity, the excess air level in the bed will be too low, and more fluidizing air must be supplied to counteract this. The incinerator's capacity is limited by its size in the sense that size limits the rate at which fluidizing air can be accepted. When fluidizing air is discharged in excess of 2,100 scfm, the metal tuyeres through which the air must pass upward into the bed are blown out.

Exhaust gases from the incinerator enter a Venturi scrubber system. Gases are scrubbed of ash and cooled and discharged to the atmosphere. Ash slurry is pumped to a Dorr-Clone wet cyclone and classifier system for dewatering. Dewatered ash is carried in a 3.06-m^3 (4-yd^3) truck to a small landfill. 2.29 m^3 (3 yd^3) of ash are produced each day. The scrubber water is recirculated back to the head of the plant. A single sample of the scrubber water showed a SS concentration of 758 ppm and a pH of 6.4.

Solids are processed in the incinerator continually, except for 32 hr on the weekends. Before the incinerator is shut down, the temperature is run up to 871°C (1,600°F). Over the next 32 hr the temperature falls to about 643°C (1,190°F). Before beginning again for the next week, preheating takes place for about 1 hr to get the temperature to 677°C (1,250°F) when feeding can start. The usual operating bed temperature is 721°C (1,330°F). Fluidizing air is normally discharged at 2,000 scfm.

Phosphorus removal has adversely affected incinerator capacity and fuel consumption because of the increased moisture content of the sludge cake fed to the incinerator. In Table 2, the incinerator feed rate and fuel consumption rate in 1970 and 1976 are indicative of the performance before and during phosphorus removal. However, part of the feed rate decrease and fuel consumption rate increase is due to the removal of the heat exchanger. The effect of phosphorus removal alone is judged to be an average decrease of approximately 121 kg (266 lb) dry solids/hr in the feed rate and an increase of 142 ℓ/t (34 gal/ton) in fuel oil consumption.

There have also been increased incinerator maintenance and repair problems with phosphorus removal due to slag formation and corrosion. On one occasion, four tuyeres blew out because several tuyeres were plugged with slag. On another occasion, a pressure build-up in the reactor signalled a problem, and a big piece of slag was found clogging the exhaust line. After this experience, the operator has done visual inspection of the exhaust line from the roof duct inspection port every 3 to 4 mo. The plant manager believes that ferric chloride is also responsible for a high rate of corrosion of the metal in ducts, especially in the elbows of the scrubber system. Gradually, these parts are being replaced with stainless steel.

Sludge Treatment and Disposal Costs

Operational Costs--

The additional costs for sludge treatment and disposal which resulted from phosphorus removal which have been estimated include the cost of chemical conditioners, polymer for the thickener, and fuel oil. The current costs of liquid sludge hauling, incinerator ash disposal, electricity for vacuum filter motors and incinerator motors, repair equipment for the incinerator, maintenance supplies for filtration and incineration combined, and labor for filtration and incineration combined, have been estimated.

In assessing the operational costs of sludge treatment and disposal at Sheboygan, the following assumptions were made:

1. All costs are shown as per ton of dry solids filtered, except for the cost of liquid sludge hauling which is per ton of solids hauled.

2. Current costs are based on a filtration rate of 9,080 kg TS/day or 3,310 t/yr (20,000 lb TS/day or 3,650 ton/yr) at a feed concentration of 7.0 percent TS; sludge hauling rate of 728 t/yr or 0.023 mil m^3/yr (803 ton/yr or 6.0 MG/yr).

3. The unit cost factors used are:

	(1977 costs)
Ferric chloride	$.0277/kg ($.0126/lb)
Conditioning polymer	$3.74/kg ($1.70/lb)
Anionic polymer	$3.13/kg ($1.42/lb)
Contractor's sludge hauling fee	$3.01/$m^3$ ($0.0114/gal)
Fuel oil	$113.34/$m^3$ (0.429/gal)
Electricity	$0.01622/kwh

4. Labor for filter and incinerator operation consists of:

160 Operator I man-hr/wk,
20 Operator II man-hr/wk

The cost breakdown is:

	Changes in costs resulting from phosphorus removal	Current costs
Ferric chloride conditioning assuming 8,000 gal/yr were used before phosphorus removal	+$1.65/t (+$1.50/ton)	$1.84/t ($1.67/ton)
Polymer conditioning	-$2.43/t (-$2.21/ton)	$4.49/t ($4.08/ton)

	Changes in costs resulting from phosphorus removal	Current costs
Thickening polymer	+\$2.20/t (+\$2.00/ton)	\$2.20/t (\$2.00/ton)
No. 2 fuel oil	+\$16.08/t (+\$14.59/ton)	\$58.65/t (\$53.20/ton)
Electrical power for chemical conditioning and vacuum filter operation	Unknown	\$2.77/t (\$2.52/ton)
Electrical power for incinerator operation	Unknown	\$3.43/t (\$3.12/ton)
Labor for filter and incinerator - 1976	Unknown	\$16.61/t (\$15.07/ton)
Liquid sludge hauling	Unknown	\$29.85/t (\$27.08/ton)
Charge for ash landfill use	None	None
Equipment for incinerator repair - 1976	Unknown	\$15.10/t (\$13.70/ton)
Equipment for vacuum filter repair - 1976	Unknown	None
Supplies for vacuum filter and incinerator maintenance - 1976	Unknown	\$4.96/t (\$4.50/ton)

Since filter yield and incinerator rates were decreased with phosphorus removal, the cost per ton of dry solids for electricity and labor for filter and incinerator operation must have increased. Because of more hours of operation of the filter and the incinerator, the cost of maintenance and repair supplies and equipment per ton of solids must also have increased. The amounts of these increases are unknown because it is not known how much additional sludge was fed to the filters as a result of phosphorus removal. Similarly, it is not known how much additional sludge was fed to the filters as a result of phosphorus removal. However, since the TS concentration of the thickened sludge was lower with phosphorus removal, the cost per ton of dry solids to haul must have increased.

Capital Costs--

The vacuum filtration and incineration system was installed in 1967. The work included installation of two vacuum filters (all new, except for the agitators) and a fluidized bed incinerator; removing a wall from an existing building; and adding a small section to the existing building to house the ash dewatering equipment. The capital cost of this project, including installation, electrical work and piping, planning and engineering fees, etc., was $738,935.18. The planning and engineering fees alone were $46,708.10. It was financed by floating a bond issue for $396,518.00 and obtaining federal aid for $388,469.00.

Summary and Conclusions

The initial impact of phosphorus removal on sludge handling appears to have been an increase in the mass of secondary sludge generated. Concurrently, the problem of intermittent floating of the thickener sludge blanket was worsened considerably, as evidenced by the large increase in the average solids concentration of the thickener overflow. Whenever thickener upsets occurred, phosphorus removal was temporarily halted; sludge feed to the thickener was temporarily halted; and secondary sludge was pumped to the primary clarifiers. Sludge was stored in the bottoms of the primary clarifiers until the thickener stabilized. Some thickening occurred in the clarifier bottoms because of this storage. Therefore, after phosphorus removal was started, the sludge pumped from the primary clarifiers to the thickener was heavier. The solids concentration of the secondary sludge was unchanged. The total volume of thickener feed sludge did not increase with phosphorus removal because of the sludge concentrating going on in the primaries.

The additional mass of sludge pumped to the thickener after phosphorus removal was started has been calculated from available data. It is 44,000 kg/mil m^3 (370 lb TS/MG) of wastewater treated, or 2.18 t (2.4 additional tons) TS/day assuming that 49,200 m^3/day (13 mgd) of wastewater are treated. Part of this increase in sludge mass was due to the recirculation of more solids in the thickener overflow. Part was due to the use of ferric chloride and polymer for phosphorus removal. Unfortunately, available data for the period we are concerned with do not permit calculation of the volume and total solids mass of the thickener overflow or the sludge pumped to the holding tanks. Therefore, we do not know what fraction of the additional 5.75 t TS/mil m^3 (2.4 tons TS/MG) which entered the thickener went out in the overflow, to the vacuum filters, or the the holding tanks.

We do know, however, that phosphorus removal resulted in a change in the dewatering characteristics of the thickened sludge. Comparing data from 6-mo periods in 1970 and 1976 which were before and after phosphorus removal began, we see that the average TS concentration of the thickened sludge decreased from 8.6 percent to 7.0 percent TS. Filter cake dryness decreased from 25.5 percent to 21.5 percent TS. The volatile content of the filter cake did not change. Filtrate SS concentration decreased. Ferric chloride conditioning became necessary to achieve a filter cake TS concentration greater than 18 percent. Previously only about 20 percent of the sludge had been conditioned with ferric chloride. Because of ferric conditioning, polymer conditioner dosage decreased. As a result, the cost for conditioning chemicals decreased by about \$0.78/t (\$0.71/ton).

Because of the higher moisture content of the filter cake, incinerator fuel oil consumption rose and the fuel oil cost rose by \$16.10/t (\$14.60/ton) of dry solids. In addition, the rate at which cake could be burned in the incinerator was lower by about 121 kg (266 lb) TS/hr. Because the incineration rate was lower, the vacuum filtration rate was lowered to match. Filter yield became only 12.8 kg TS/m^2/hr (2.62 lb TS/ft^2/hr). A higher filter yield is possible, but at the expense of cake dryness. It is difficult to consider filter yield independently of the incineration capacity in this situation.

In response to the adverse effect of phosphorus removal on thickener sludge blanket stability and overflow solids concentration, efforts were made to find a method of adding polymer to the thickener which would improve the situation. A really successful method was not found until January of 1977. Before this, in June of 1975, the thickener overflow was rerouted to go to the primary clarifier influent instead of the wet well ahead of the trickling filters. Poor thickener overflow quality had an adverse effect on plant BOD and SS removal and plant effluent quality. BOD and SS removal are so dependent upon influent flow and strength, however, that it has not been possible to present data showing the effect of the thickener overflow. Nevertheless, improvement in plant BOD and SS removal can be considered to be benefits derived from the addition of polymer to the thickener. Another benefit, better phosphorus removal, occurred because ferric and polymer addition no longer had to be halted frequently for thickener upsets. The costs of polymer addition, on the other hand, are the cost of the polymer itself, which is \$2.21/t (\$2.00/ton) of dry solids, and the cost of disposing of the additional sludge solids which are generated as a result of polymer usage.

CASE STUDY E: COLDWATER, MICHIGAN

Introduction

The Coldwater, Michigan treatment plant provides an example of the primary addition of ferric chloride for phosphorus removal, with and without the use of anionic polymer. The major components of the Coldwater treatment plant are trickling filters, pressure sand filters, anaerobic digesters, and sludge drying beds.

Wastewater influent to the Coldwater plant is delivered by separate sanitary sewers, although some older portions of the collection system do experience groundwater infiltration. Less than 5 percent of the wastewater flow is from industrial sources. One of the industrial sources is a slaughterhouse, which contributes 379 m^3/day (0.1 mgd) with BOD concentrations between 200 and 800 mg/ℓ. The other industrial source is a rendering plant which contributes 1,140 m^3/day (0.3 mgd) with BOD concentrations as high as 2,700 mg/ℓ.

Three years of treatment plant operational data have been selected for study as follows:

- 1971 - Prior to initiation of chemical addition for phosphorus removal.
- 1973 - With chemical addition of both ferric chloride and anionic polymer.
- 1976 - With chemical addition of ferric chloride only.

Table C-9 presents the average influent and effluent wastewater characteristics and plant removal efficiencies for each time period.

TABLE C-9. INFLUENT AND EFFLUENT WASTEWATER CHARACTERISTICS AND REMOVAL EFFICIENCIES, COLDWATER, MICHIGAN

		1971	1973	1976
FLOW m^3/d (mgd)		4,660 (1.23)	7,310 (1.93)	9,200 (2.43)
SS (mg/ℓ) -	influent	156	114	168
	effluent	20	7	15
	% removal	87	93	91
BOD (mg/ℓ)	influent	297	289	253
	effluent	30	21	26
	% removal	90	93	90
TP (mg/ℓ)	influent	---	6.2	5.4
	effluent	---	1.2	0.7
	% removal	---	80	86

History

Modifications to the plant which have affected sludge production and characteristics are detailed below:

1952 Original secondary wastewater treatment plant constructed. Treatment units consisted of a grit chamber, two primary clarifiers, two trickling filters. one final clarifier (identical to the

	primary clarifiers), and chlorination. Sludge from the final clarifier was returned to the head of the primary clarifiers, settled, and subsequently removed from the bottom of the primary clarifiers along with the settled wastewater solids. This sludge was pumped to the primary sludge digester where it was heated and mixed and then transferred to the secondary digester. Sludge in the secondary digester was allowed to settle without heating or mixing, while the supernatant was returned to the flocculation channel ahead of the primary clarifiers. Sludge from the secondary digester was eventually transferred to one of the sludge drying beds. The dry sludge cake was disposed of at a landfill site.
1971	Construction of tertiary sand filters began.
April 1972	Plant flow increased by 946 mg^3/d (0.25 mgd) when new trunk sewer from "State Home" connected.
July 1972	Two new final clarifiers on line and one old secondary clarifier converted to perform primary clarification in parallel with two existing primary clarifiers; third trickling filter also on line at this time.
Fall 1972	Existing five sludge drying beds paved and seven new paved sludge drying beds constructed; new above-ground, unheated holding digester added; new gas lifter system installed in the existing primary digester.
Sept. 1972	Tertiary sand filters put into operation.
Dec. 1972	Chemical addition for phosphorus removal began; ferric chloride initially added to the aerated grit chamber, while anionic polymer was added to the head of the aerated flocculation at the flash mixer channel.
1976	Anionic polymer addition suspended.
1977	Anionic polymer addition resumed using liquid polymer instead of dry polymer.

Chemical Addition for Phosphorus Removal

Liquid ferric chloride is added to the head of the grit chamber. Anionic polymer (Haviland Chemical, Poly-floc M-P) is then added at the flash mixer at the head of the flocculation channel. Approximately 9.4 min contact time at 7,570 m^3/day (2 mgd) plant flow is provided in the flocculation channel. The chemical dosage is selected to achieve 90 percent phosphorus removal. The average ferric chloride dosage was 40 mg/ℓ dry $FeCl_3$ until 1977, when it was raised to 75 mg/ℓ. The average dosage of anionic polymer is 4 mg/ℓ.

General Description of Wastewater Treatment Operations Affecting Sludge

A flow diagram for the wastewater and sludge treatment operations is presented in Figure C-6. Table C-10 gives a summary of major equipment at the Coldwater plant utilized for wastewater treatment and sludge handling. In addition to the specific modifications to the wastewater treatment plant listed in Table C-9, there are some seasonal variations in plant operations and plant influent characteristics.

Generally, two trickling filters are operated to provide secondary treatment. The third trickling filter is not operated unless it can be run for at least three months. It has been frequently necessary, during the winter months, to use only one trickling filter. This serves to increase flow rates through the filter, thereby avoiding freeze-ups.

A variation in plant influent characteristics is due to the dumping of the concentrated wastes pumped out of septic tanks. These dumpings occur with highest frequency during the summer months, causing brief but significant increases in plant influent characteristics. Specific data describing these influents were not available; however, plant personnel indicated that the quantities of wastes delivered to the plant have remained relatively constant throughout the study periods.

A flow diagram and materials balance of the Coldwater plant are shown in Figure C-7. The data presented are based on information contained in the monthly reports submitted by the plant to the state of Michigan, during the years selected for study (viz. - 1971, 1973, and 1976). From the figure, it can be seen that wastewater solids information is presented in terms of suspended solids (SS), while sludge and sidestream information is given as total solids (TS). In spite of the inconsistency of the units of expression, all available solids information has been presented to qualitatively substantiate the observed impacts of chemical additions for phosphorus removal on plant solids handling operations. The remainder of this section briefly describes each of the wastewater treatment unit operations which have significant impacts on sludge.

Flocculation Channel--

Figure C-8 is a flow diagram of the flocculation channel. The flocculation channel receives raw influent wastewater after degritting, chemical addition for phosphorus removal, and pressure sand filter backwash water return have taken place.

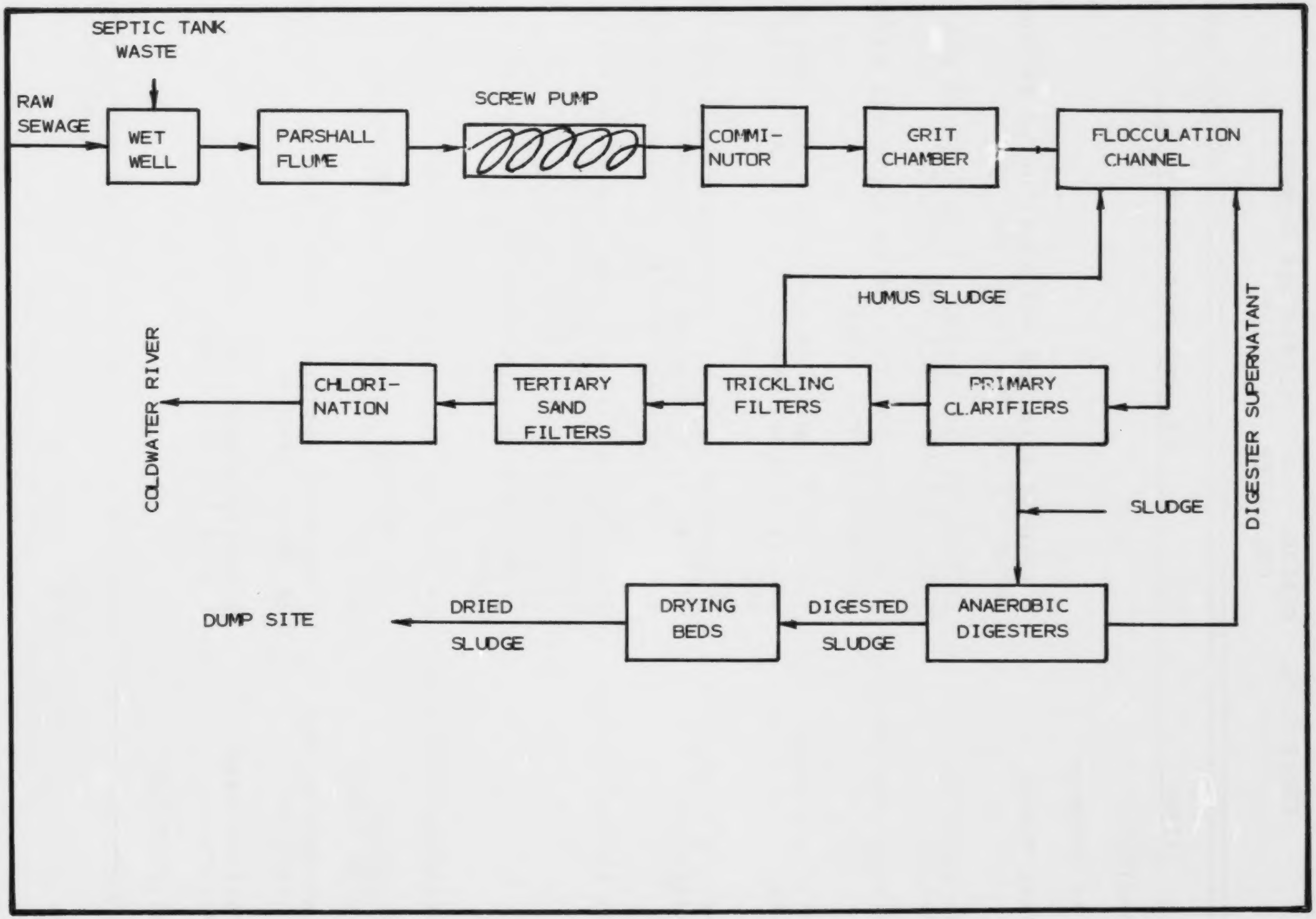

SEPTIC TANK WASTE
RAW SEWAGE
WET WELL
PARSHALL FLUME
SCREW PUMP
COMMI-NUTOR
GRIT CHAMBER
FLOCCULATION CHANNEL
HUMUS SLUDGE
DIGESTER SUPERNATANT
COLDWATER RIVER
CHLORI-NATION
TERTIARY SAND FILTERS
TRICKLING FILTERS
PRIMARY CLARIFIERS
SLUDGE
DUMP SITE
DRIED SLUDGE
DRYING BEDS
DIGESTED SLUDGE
ANAEROBIC DIGESTERS

Figure C-6. Coldwater, Michigan, wastewater treatment plant flow diagram.

TABLE C-10. GENERAL PLANT DESCRIPTION SUMMARY, COLDWATER, MICHIGAN

Item	No. and Description
Comminutor	
Grit Chamber	
Flash Mixer	1 with 91.4 cm (36 in) impeller at 84 rpm; motor - 5 hp, 30, 60 hz, 460 v.
Flocculation Channel	27.7 m (91 ft) length x 1.30 m (4.25 ft) width x 1.37 m (4.5 ft) height
Primary Clarifiers	3 at 27.7 m (91 ft) x 4.88 m (16 ft) width x 2.74 m (9 ft) height
Primary Sludge Pumps	3 at 34.1 ℓ/min (40 gal/min) - one pump use at a time; motors - 3 hp, 30, 60 hz, 460 v.
Primary Sludge Collectors	3 pairs of vertical chains with wooden flights at 3.05 m (10 ft) intervals; motors - 2 at 1/2 hp, 30, 60 hz, 460 v.
Trickling Filters	3 at 35.1 m (115 ft) dia. x 2.44 m (8 ft) side wall depth (SWD) rack-filled (pit run) field stone, 3.81 to 10.2 cm (1.5 to 4 in dia). Flow sprayed continuously from fixed nozzles rotating at 2 to 20 rpm
Secondary Clarifiers	19.8 m (65 ft) dia. x 3.05 m (10 ft) SWD
Secondary Sludge Pumps	2 at 75 rpm; motors - 2 hp, 30, 60 hz, 460 v.
Secondary Sludge Collectors	2 rotating collector arms; motors - 3/4 hp, 30, 60 hz, 460 v.
Pressure Sand Filters	4 multiple media at 1.83 m (6 ft) dia. x 4.57 (15 ft) length
Chlorination	

(continued)

TABLE C-10 (continued)

Item	No. and Description
Primary Digester	958,000 ℓ (253,000 gal) capacity with 13.7 m (45 ft) dia. x 5.49 m (18 ft) SWD; fixed concrete dome and 3 gas lift eductor tube encased by an internal water heating jacket assembly; eductor tube capacity - 41,600 ℓ (11,000 gal)/min total; water
Primary Digester	jacket heat exchange capacity 500,000 Btu/hr; motor - 10 hp, 3Ø, 60 hz, 460 v; boiler converter pump - 2 hp, 3Ø, 60 hz, 460 v.
Secondary Digester	958,000 ℓ (253,000 gal) capacity at 13.7 m (45 ft) dia. x 5.49 m (18 ft) SWD with fixed concrete dome
Holding Digester	142,000 ℓ (374,000 gal) capacity at 16.8 m (55 ft) dia. x 5.79 m (19 ft) SWD; with fixed steel dome
Sludge Drying Beds	Concrete paved; 12 at 32.9 m (108 ft) length x 7.62 m (25 ft) width x 3.81 m (125 ft) height
Sludge Lagoon	700,000 ℓ (185,000 gal) capacity at 30.5 m (100 ft) length x 15.2 m (50 ft) width x 1.52 m (5 ft) depth

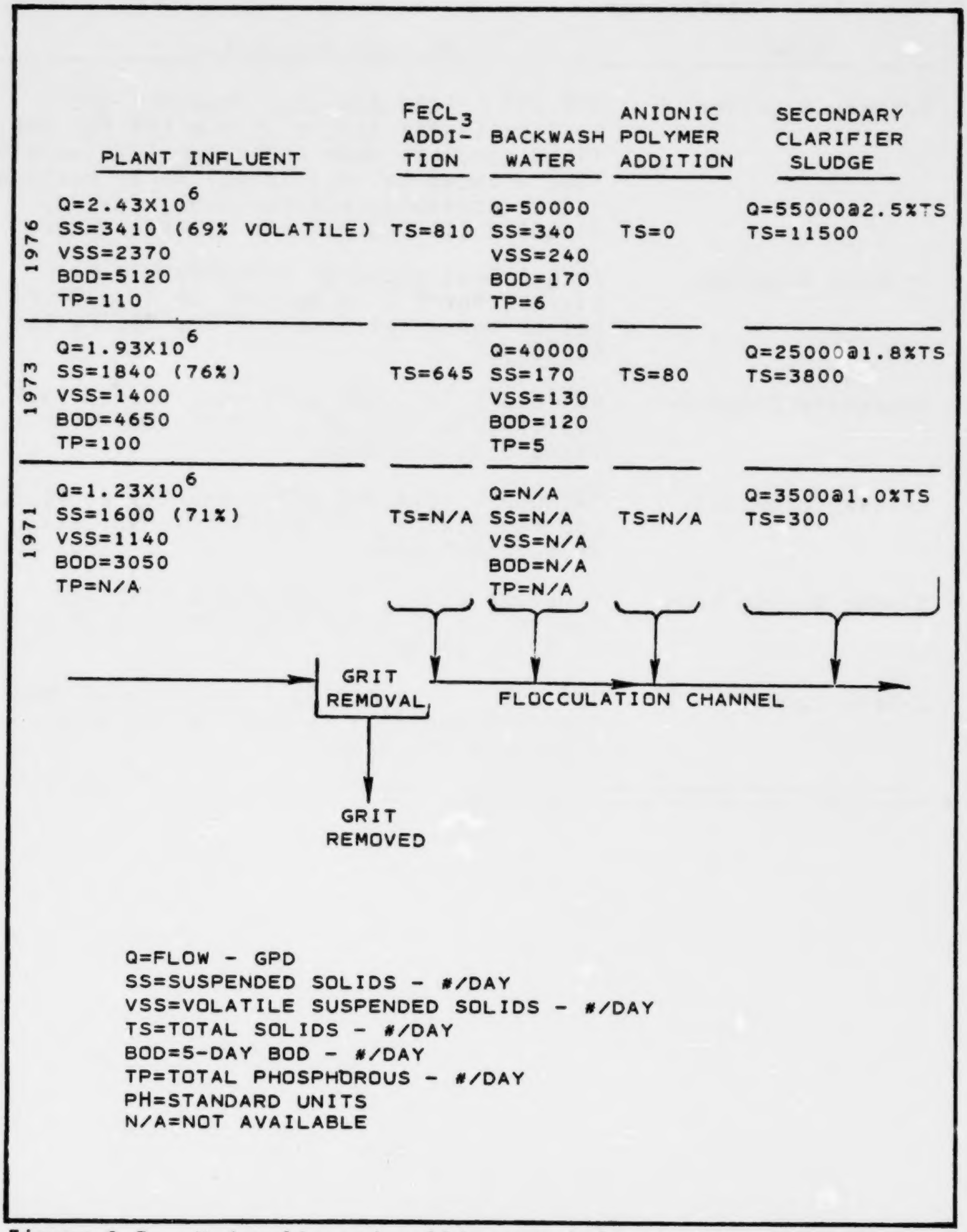

PLANT INFLUENT
$FeCl_3$ ADDITION
BACKWASH WATER
ANIONIC POLYMER ADDITION
SECONDARY CLARIFIER SLUDGE
1976
$Q=2.43X10^6$
SS=3410 (69% VOLATILE)
VSS=2370
BOD=5120
TP=110
TS=810
Q=50000
SS=340
VSS=240
BOD=170
TP=6
TS=0
Q=55000@2.5%TS
TS=11500
1973
$Q=1.93X10^6$
SS=1840 (76%)
VSS=1400
BOD=4650
TP=100
TS=645
Q=40000
SS=170
VSS=130
BOD=120
TP=5
TS=80
Q=25000@1.8%TS
TS=3800
1971
$Q=1.23X10^6$
SS=1600 (71%)
VSS=1140
BOD=3050
TP=N/A
TS=N/A
Q=N/A
SS=N/A
VSS=N/A
BOD=N/A
TP=N/A
TS=N/A
Q=3500@1.0%TS
TS=300
GRIT REMOVAL
FLOCCULATION CHANNEL
GRIT REMOVED
Q=FLOW - GPD
SS=SUSPENDED SOLIDS - #/DAY
VSS=VOLATILE SUSPENDED SOLIDS - #/DAY
TS=TOTAL SOLIDS - #/DAY
BOD=5-DAY BOD - #/DAY
TP=TOTAL PHOSPHOROUS - #/DAY
PH=STANDARD UNITS
N/A=NOT AVAILABLE

Figure C-7. Hydraulic and solids mass balances for wastewater and sludge treatment operations, Coldwater, Michigan.

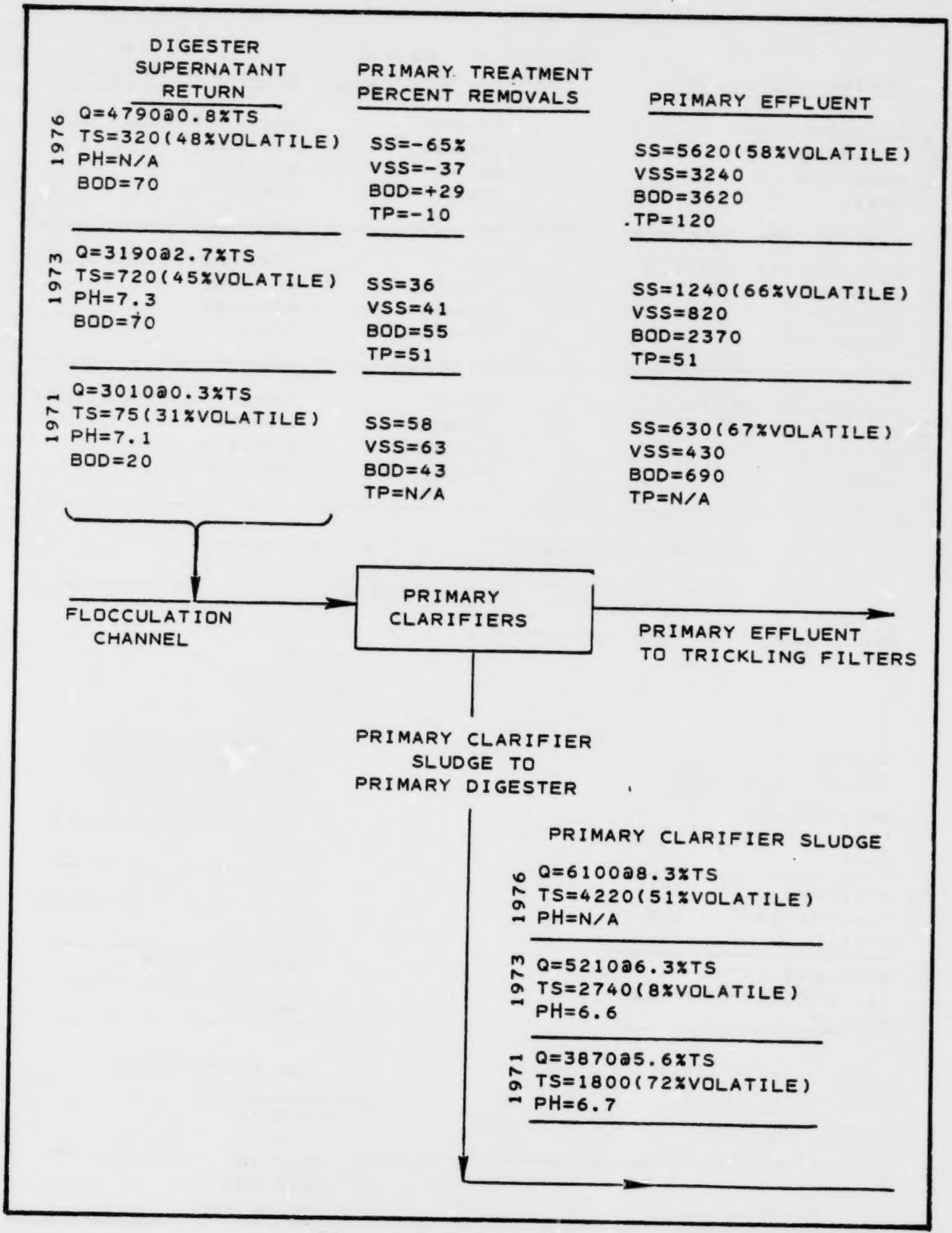
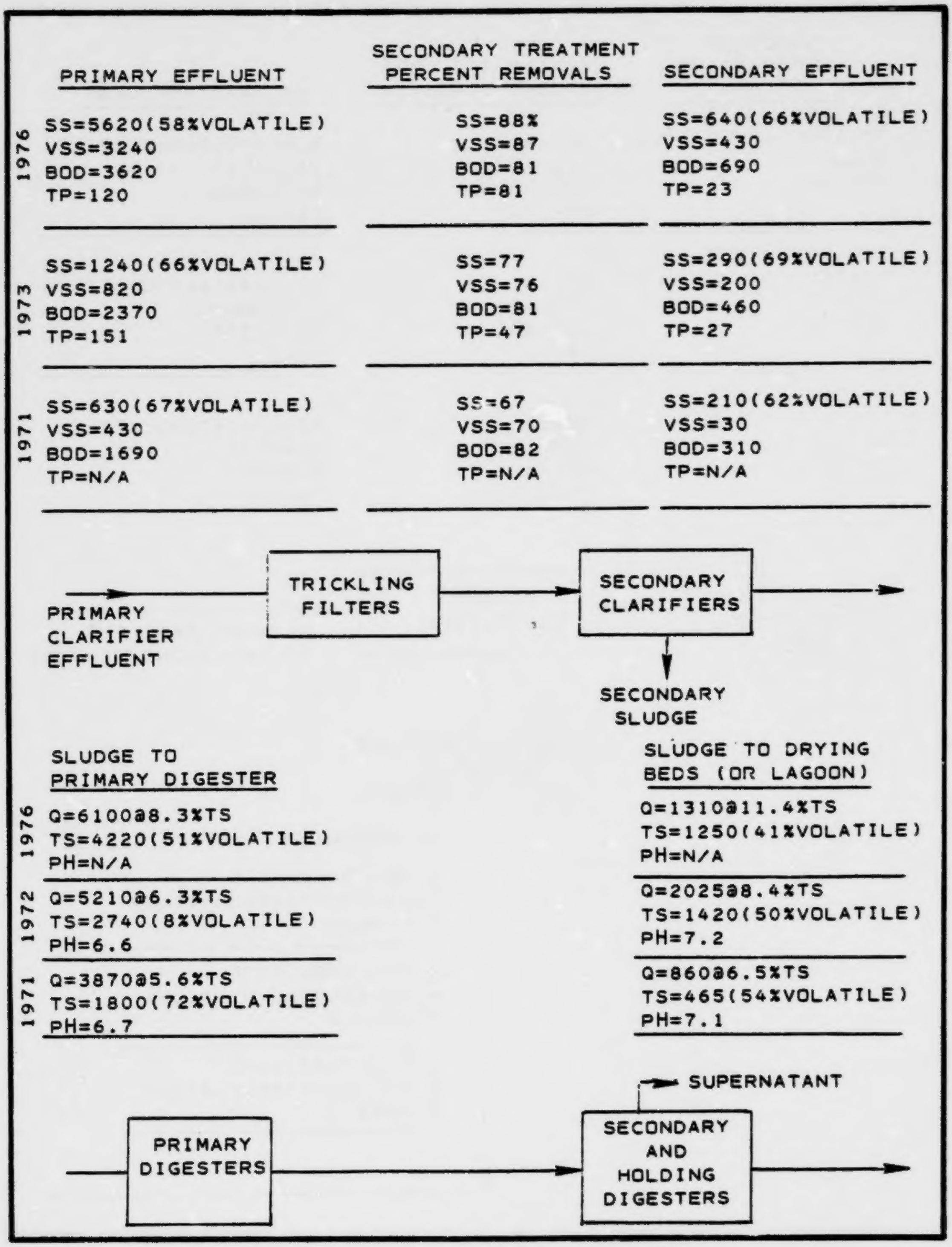

DIGESTER SUPERNATANT RETURN
1976
Q=4790@0.8%TS
TS=320(48%VOLATILE)
PH=N/A
BOD=70
1973
Q=3190@2.7%TS
TS=720(45%VOLATILE)
PH=7.3
BOD=70
1971
Q=3010@0.3%TS
TS=75(31%VOLATILE)
PH=7.1
BOD=20
PRIMARY TREATMENT PERCENT REMOVALS
SS=-65%
VSS=-37
BOD=+29
TP=-10
SS=36
VSS=41
BOD=55
TP=51
SS=58
VSS=63
BOD=43
TP=N/A
PRIMARY EFFLUENT
SS=5620(58%VOLATILE)
VSS=3240
BOD=3620
TP=120
SS=1240(66%VOLATILE)
VSS=820
BOD=2370
TP=51
SS=630(67%VOLATILE)
VSS=430
BOD=690
TP=N/A
FLOCCULATION CHANNEL
PRIMARY CLARIFIERS
PRIMARY EFFLUENT TO TRICKLING FILTERS
PRIMARY CLARIFIER SLUDGE TO PRIMARY DIGESTER
PRIMARY CLARIFIER SLUDGE
1976
Q=6100@8.3%TS
TS=4220(51%VOLATILE)
PH=N/A
1973
Q=5210@6.3%TS
TS=2740(8%VOLATILE)
PH=6.6
1971
Q=3870@5.6%TS
TS=1800(72%VOLATILE)
PH=6.7

Figure C-7 (continued)



TRICKLING FILTERS
PRIMARY CLARIFIER EFFLUENT
SECONDARY CLARIFIERS
SECONDARY SLUDGE
SUPERNATANT
PRIMARY DIGESTERS
SECONDARY AND HOLDING DIGESTERS

	PRIMARY EFFLUENT	SECONDARY TREATMENT PERCENT REMOVALS	SECONDARY EFFLUENT
1976	SS=5620(58%VOLATILE) VSS=3240 BOD=3620 TP=120	SS=88% VSS=87 BOD=81 TP=81	SS=640(66%VOLATILE) VSS=430 BOD=690 TP=23
1973	SS=1240(66%VOLATILE) VSS=820 BOD=2370 TP=151	SS=77 VSS=76 BOD=81 TP=47	SS=290(69%VOLATILE) VSS=200 BOD=460 TP=27
1971	SS=630(67%VOLATILE) VSS=430 BOD=1690 TP=N/A	SS=67 VSS=70 BOD=82 TP=N/A	SS=210(62%VOLATILE) VSS=30 BOD=310 TP=N/A

	SLUDGE TO PRIMARY DIGESTER	SLUDGE TO DRYING BEDS (OR LAGOON)
1976	Q=6100@8.3%TS TS=4220(51%VOLATILE) PH=N/A	Q=1310@11.4%TS TS=1250(41%VOLATILE) PH=N/A
1972	Q=5210@6.3%TS TS=2740(8%VOLATILE) PH=6.6	Q=2025@8.4%TS TS=1420(50%VOLATILE) PH=7.2
1971	Q=3870@5.6%TS TS=1800(72%VOLATILE) PH=6.7	Q=860@6.5%TS TS=465(54%VOLATILE) PH=7.1

Figure C-7 (continued)

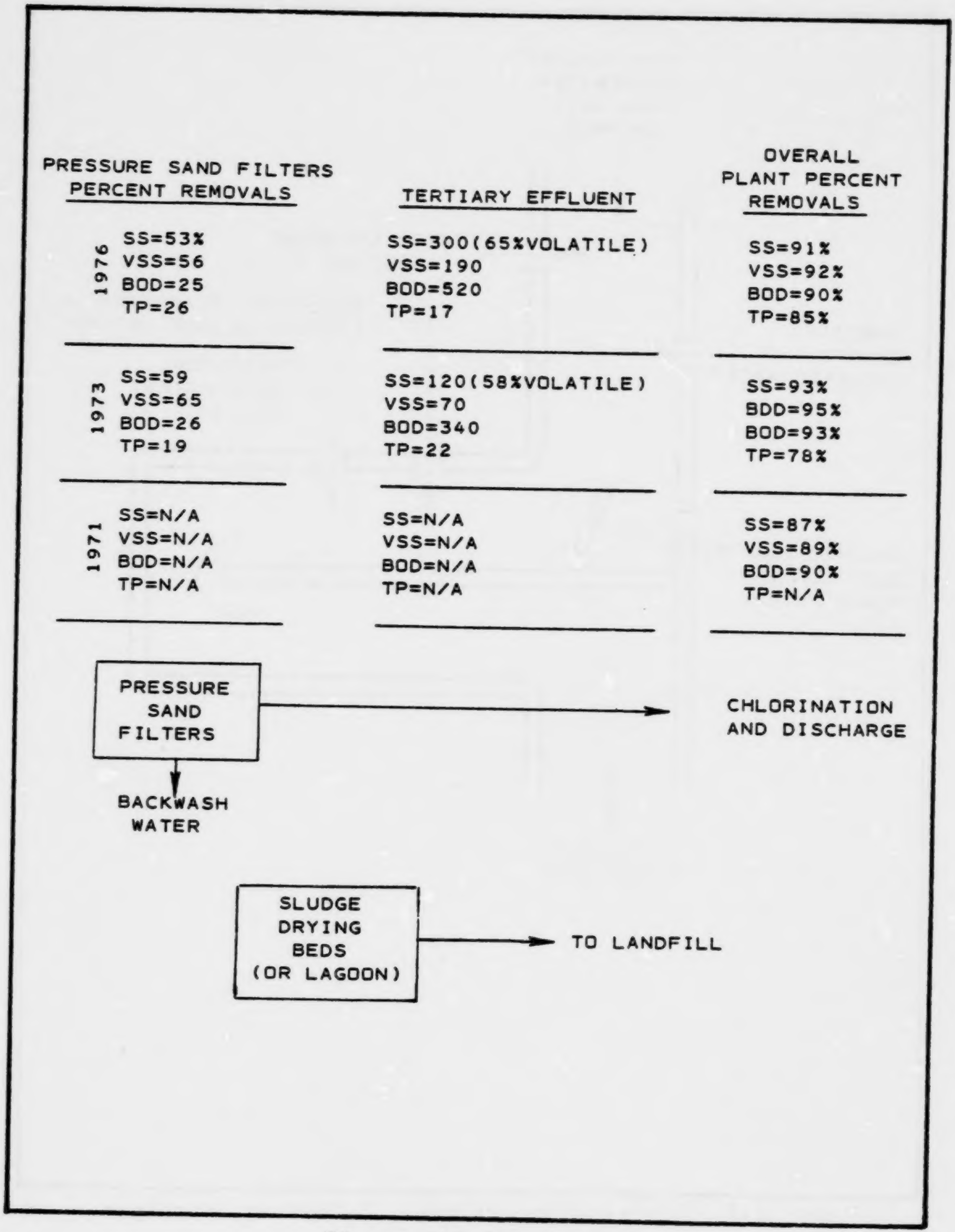

PRESSURE SAND FILTERS PERCENT REMOVALS
TERTIARY EFFLUENT
OVERALL PLANT PERCENT REMOVALS
1976
SS=53%
VSS=56
BOD=25
TP=26
SS=300(65%VOLATILE)
VSS=190
BOD=520
TP=17
SS=91%
VSS=92%
BOD=90%
TP=85%
1973
SS=59
VSS=65
BOD=26
TP=19
SS=120(58%VOLATILE)
VSS=70
BOD=340
TP=22
SS=93%
BDD=95%
BOD=93%
TP=78%
1971
SS=N/A
VSS=N/A
BOD=N/A
TP=N/A
SS=N/A
VSS=N/A
BOD=N/A
TP=N/A
SS=87%
VSS=89%
BOD=90%
TP=N/A
PRESSURE SAND FILTERS
CHLORINATION AND DISCHARGE
BACKWASH WATER
SLUDGE DRYING BEDS (OR LAGOON)
TO LANDFILL

Figure C-7 (continued)

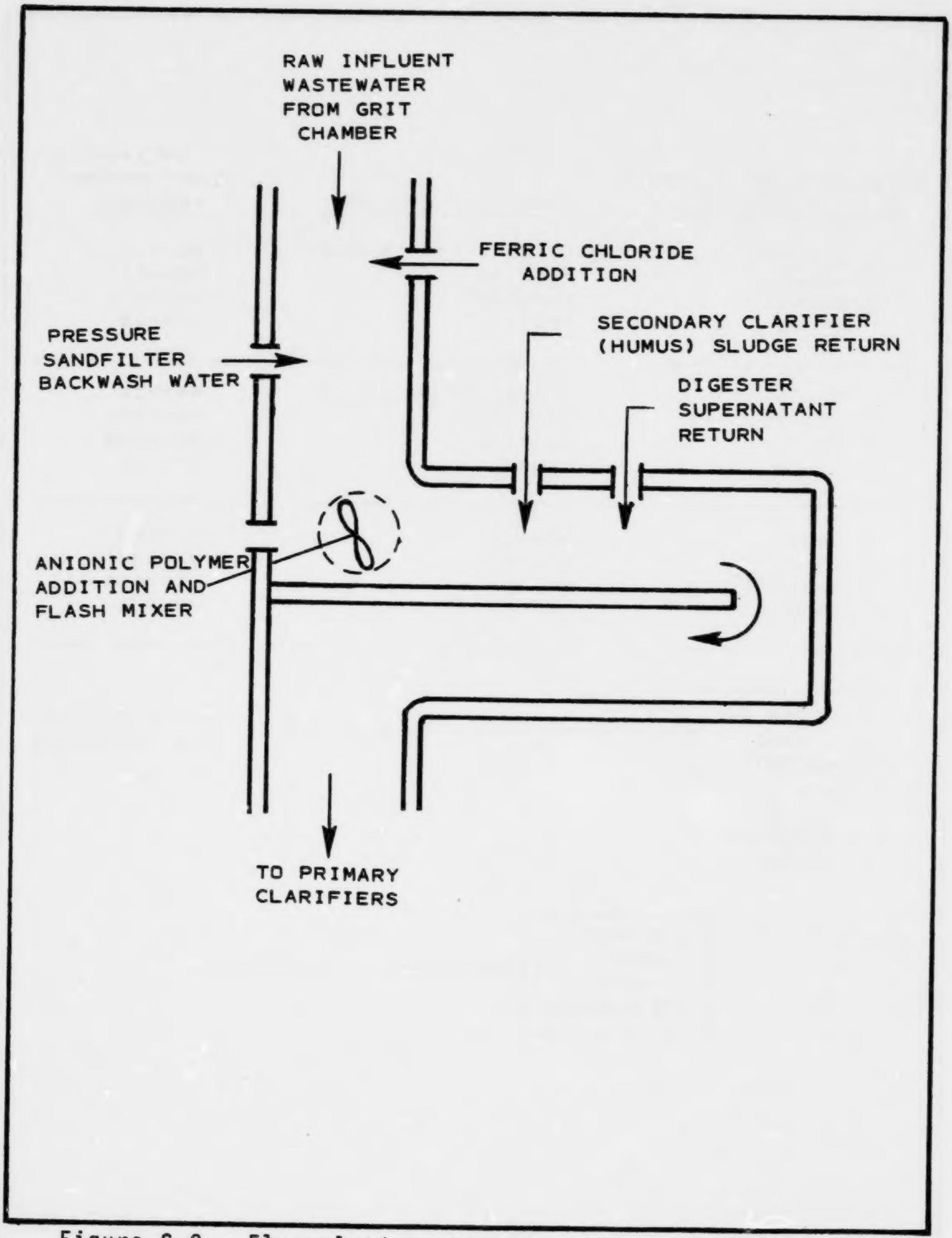

RAW INFLUENT WASTEWATER FROM GRIT CHAMBER
FERRIC CHLORIDE ADDITION
PRESSURE SANDFILTER BACKWASH WATER
SECONDARY CLARIFIER (HUMUS) SLUDGE RETURN
DIGESTER SUPERNATANT RETURN
ANIONIC POLYMER ADDITION AND FLASH MIXER
TO PRIMARY CLARIFIERS

Figure C-8. Flocculation channel, Coldwater, Michigan.

Digester supernatant and secondary clarifier sludge enter the flocculation channel approximately 6.10 m (20 ft) from the flash mixer at the head of the channel. Ferric chloride solutions are added to the grit chamber, which is ahead of the flash mixer, and anionic polymer addition takes place at the flash mixer. As previously mentioned, polymer addition began in 1972, was discontinued in 1976, and was reinstituted in 1977.

With the advent of chemical addition for phosphorus removal, the amount of solids entering the flocculation channel in the pressure sand filter backwash water has increased steadily. Solids contained in the supernatant also increased considerably after phosphorus removal began, although somewhat less significantly after polymer addition was suspended. The most significant increase, however, has been in sludge solids returned to the flocculation channel from the secondary clarifiers. The combined effect of these increases has caused no problems in the flocculation channel but has resulted in huge increases in solids loadings on the primary clarifiers.

Primary Clarifiers--

The primary clarifiers receive the raw influent wastewater and the returned sidestreams from the flocculation channel. Sludge is collected and removed from the bottom of each clarifier 3 times/day, 7 days/wk. During the intervals between sludge collection and pumping, a sludge blanket as thick as two feet builds up on the bottoms of the primary clarifiers. Settled sludge is pushed to the sludge hopper in each clarifier by wooden flights attached to two parallel endless chains. The sludge collectors are turned on a half hour prior to sludge pumping from the clarifier and are operated until sludge pumping is completed.

Sludge collected in the sludge hoppers is pumped by the two primary sludge pumps to the primary digester from one clarifier at a time. The sludge solids concentration is approximately 10 percent TS when pumping begins. Pumping continues until the solids concentration is reduced to roughly 3.5 percent TS.

Since the initiation of chemical addition for phosphorus removal, the amount of solids influent to the primary clarifiers and both the quantity and total solids concentration of the sludge pumped from the primary clarifier to the primary digester has increased. Concurrently, the volatile fraction of these solids has steadily declined. Due to variations in plant influent suspended solids concentrations and flow rates (previously characterized in Table C-9), the impacts of phosphorus removal on sludge generation rates cannot be adequately compared on the basis of pounds of primary sludge solids to the digester per mg of plant influent. Instead, Table C-11 relates the mass of total solids to the digester to the mass of suspended solids in the plant influent.

TABLE C-11. SLUDGE PUMPED FROM PRIMARY CLARIFIERS TO PRIMARY DIGESTER, COLDWATER, MICHIGAN

Year	ℓ/day (gal/day)	TS (%)	Volatile (%)	kg(lb)TS/kg (lb) SS plant influent	kg(lb)VS/kg(lb) VSS plant influent	kg(lb)FS/kg(lb) FSS plant influent
1976	23,100 (6,100)	8.3	51	0.56 (1.24)	0.41 (0.91)	0.90 (2.00)
1973	19,700 (5,210)	6.3	58	0.67 (1.48)	0.51 (1.13)	1.18 (2.61)
1971	14,600 (3,870)	5.6	72	0.51 (1.12)	0.52 (1.12)	0.49 (1.09)

On this basis, it can be seen that the mass of volatile solids to the digester per pound volatile suspended solids in the plant influent remained roughly the same after the addition of both ferric chloride and polymer began and decreased slightly when only ferric chloride was added. On the other hand, the mass of total solids to the digester per pound suspended solids in the plant influent showed a significant increase when both polymer and ferric chloride were added and a moderate increase when only ferric chloride was added. These increases are apparently attributable to the increases in the nonvolatile (fixed) portion of the sludge solids, due to chemical additions for phosphorus removal.

Trickling Filters--

A notable impact of chemical addition for phosphorus removal on the trickling filters at the Coldwater plant has been the reddish-brown discoloration of the trickling filter structure and media. Although BOD loadings to the trickling filters have increased significantly, percent removals of BOD have remained constant. Higher BOD concentrations are thus found in the trickling filter effluent.

Secondary Clarifiers--

Trickling filter effluent flows by gravity to a division well which distributes the flow to the two secondary clarifiers. Each clarifier is equipped with a pair of rotating collector arms equipped with squeegees to continuously move settled sludge to a sludge hopper at the bottom of the tank. One of the collector arms is equipped with a traveling scum collector which moves the collected grease and scum to a fixed scum box assembly. Scum then enters a scum well along with the secondary sludge and is pumped to the flocculation channel at the head of the primary clarifier.

Sludge which has settled to the bottoms of the secondary clarifiers is pumped twice a day to the flocculation channel. Sludge is pumped from one final clarifier for the same amount of time. Both pumps are used to develop enough suction to prevent clogging of the pumps. When sludge pumping begins, the sludge solids concentration is approximately 5 percent TS. Sludge pumping continues until the liquid delivered to the flocculation channel no longer has a black color, which corresponds roughly to a solids concentration of 0.1 percent TS.

Since the beginning of chemical addition for phosphorus removal, both the secondary sludge volume and total solids concentration have increased significantly. This has caused a huge increase in the TS mass pumped to the flocculation channel. This increase is the most notable change in plant operations since chemical additions for phosphorus removal were started.

Pressure Sand Filters--

Secondary effluent flows to the pressure sand filters where approximately half of the remaining suspended solids and one fourth of the remaining BOD and phosphorus are removed. Backwash water from the pressure sand filters is discharged to a holding basin where some solids settling takes place. Backwash water in excess of basin capacity overflows the holding basin weir and enters the flocculation channel. The quantities and characteristics of the backwash water were shown in Figure C-10.

Detailed Description of Sludge Treatment and Disposal Operations

Primary Digester--

Sludge enters the primary digester 3 times/day, 7 days/wk from the primary clarifiers. The primary digester is heated and mixed. It is generally maintained at a temperature between 29 and 35°C (85 and 95°F). Heating and mixing is intermittent, taking place for at least 16 hr/day, with longer periods during the fall and winter months. Mixing of the digester takes place by compressing the collected gas from the destruction of volatile materials and discharging it through three gas eductor tubes enclosed in a hot water jacket in the center of the digester. The boiler used to heat the jacket water can be operated on natural gas or methane digester gas.

The primary digester is maintained full at all times, except when sludge is withdrawn and transferred to the secondary (or, on occasion, holding) digester. During a digester sludge transfer, between 114 and 265 m^3 (30,000 and 70,000 gal) are pumped during the day shift over a period of three or more days. This is done in an effort to maximize operator efficiency, since the positions of a considerable number of valves must be changed each time sludge is transferred to or from a different location (primary, secondary, or holding digester, or sludge drying beds). After a sludge withdrawal from the primary digester, sludge transfer does not take place again until the digester has been refilled. This is because the primary digester is the only digester which is heated or mixed and is supposed to account for the majority of the volatile destruction taking place in all the digesters.

Secondary Digester--

Prior to the plant modifications in 1972, the secondary digester was virtually identical to the primary digester. At present, the secondary digester sludge is neither heated nor mixed, although the side walls of the digester are surrounded by soil. Consequently, the sludge which is transferred from the primary to the secondary digester retains much of its heat, allowing digestion and destruction of volatile materials to continue while solids separation takes place. Supernatant from this digester flows by gravity to the scum well ahead of the primary clarifiers. Although minor adjustment of the supernatant overflow pipe height is possible, the quality of the supernatant returned is largely dependent on the need for providing additional space in the secondary digester by allowing more supernatant of less than optimum solids concentration to be returned to the primary clarifier. This effect appears to have been most pronounced during 1973, when an estimated 25 percent of the total solids sent to the primary digester were returned to the flocculation channel as supernatant.

Holding Digester--

The holding digester was installed along with the other plant modifications of 1972. This digester is situated aboveground and is neither heated nor mixed. As a result, only a small amount of volatile solids destruction takes place in this digester, particularly during the winter months.

Sludge from the secondary digester (and occasionally from the primary digester) is transferred to the holding digester for storage until weather conditions or space allow transfer of sludge to the sludge drying beds. Thus, digester sludge handling operations are directed at having as little sludge as is possible in the holding digester at the beginning of autumn to maximize the amount of digester volume available for storage of sludge generated during the winter months when the sludge drying beds are essentially inoperative. Due to the format of the data submitted by the treatment plant to the state of Michigan, characteristics and quantities of sludge stored in the secondary and holding digesters are indistinguishable.

Digester Performance--

It is difficult to assess the impact of chemical addition for phosphorus removal on the individual digesters due to the limitation in available data. No measurements of the volume of supernatant returned to the flocculation channel from the secondary or holding digester are available. Therefore, it has been assumed that the volume of supernatant returned is equal to the difference between the volume of sludge pumped to the primary digester minus the volume of sludge sent to the sludge drying beds (and lagoons).

Since the start-up of chemical addition for phosphorus removal, the percent total solids of sludge entering and leaving the digester has increased. A steady decrease in the fraction of volatile solids in the sludge fed to the digesters has been observed. But the mass of VS fed to the primary digester/day increased as follows: 590 kg VS/day (1,300 lb VS/day) in 1971; 722 kg VS/day (1,590 lb VS/day) in 1973; and 976 kg VS/day (2,150 lb VS/day) in 1976. The amount of digester gas produced increased from roughly 5,040 m^3/mo (180,000 ft^3/mo) before phosphorus removal to more than 11,200 m^3/mo (400,000 ft^3/mo) after phosphorus removal.

The sludge transferred from the secondary and holding digesters to the sludge drying beds (or the lagoons), also reflects the trend of decreasing volatile solids fraction with increasing total solids concentration. Table C-12 indicates that the increase in the mass of TS transferred from the digester to the sludge drying beds, per pound of suspended solids in the plant influent, was due to a large extent to the increase in the nonvolatile (fixed) portion of the solids. Once again, this effect appears to have been most pronounced when both ferric chloride and polymer were added to the flocculation channel. Furthermore, less volatile destruction took place in the digestion process overall during this period. This resulted in a more than doubled mass of volatile solids in the sludge transferred to the drying beds, per pound of volatile suspended solids in the plant influent, as shown in Table C-12. This is in spite of the fact that, as

TABLE C-12. SLUDGE TRANSFERRED FROM DIGESTERS TO DRYING BEDS (OR LAGOON), COLDWATER, MICHIGAN

Year	ℓ/day	gal/day	TS (%)	Volatile (%)	kg TS kg(lb) SS Plant Influent	kg VS kg(lb) VSS Plant Influent	kg FS kg(lb) FSS Plant Influent
1976	5,960	1,310	11.4	41	0.167 (0.37)	0.099 (0.22)	0.322 (0.71)
1973	7,660	2,025	8.4	50	0.349 (0.77)	0.231 (0.51)	0.730 (1.61)
1971	3,870	860	6.5	54	0.131 (0.29)	0.099 (0.22)	0.213 (0.47)

was shown in Table C-11, the sludge fed to the primary digester contained nearly equal amounts of volatile solids when compared on the same basis.

Sludge Drying Beds--

Sludge is transferred to the sludge drying beds from the holding digester (or occasionally the secondary digester) at a rate largely dependent on prevailing weather conditions. Once each bed is filled, sludge is not removed until it is determined to be sufficiently dry, roughly 6.5 percent moisture by weight. Depending on weather conditions, each bed can be filled three to four times per year. The sludge is transferred to the beds in the fall, generally remains through the cold winter months, and is removed in the spring. As shown in Figure C-7, the TS concentration of the sludge transferred to the drying beds has increased considerably, while the volatile fraction of the solids has decreased since the addition of phosphorus removal chemicals. As a consequence of these trends, combined with the effects of paving the former drying beds, the sludge now takes 30 to 50 percent longer to reach the dryness desired for removal and disposal.

Sludge Lagoon--

The sludge drying lagoon was excavated shortly after chemical addition for phosphorus removal was initiated to provide additional sludge drying space. 186 s was necessitated by the longer sludge drying times, as well as the greater-than-anticipated quantities of digested sludge generated. Sludge is sent to the lagoon on a sporadic basis at times when the drying beds are full and there is a need to make room in the digesters. No formal records have been kept as to the characteristics of the sludge in the lagoon.

Sludge Treatment and Disposal Costs

Operating and Maintenance Costs--

The annual O&M costs for the Coldwater plant are based on the following unit costs:

Item	Unit Cost
Fe Cl_3	\$0.11/kg (\$0.05/lb)
Natural Gas	\$0.0215/m^2 (\$0.002/ft^3)
Electricity	\$0.0368/KWH
Labor (including supplies and benefits)	\$7.75/hr

O&M costs for the 1976-77 fiscal year were as follows:

Item	$/Yr
Operator and Superintendent Labor	$ 45,000
Operating Supplies	40,000
Electricity	4,160
Natural Gas	3,660
Water	7,840
Maintenance and Repairs	25,000
Total	$125,660
Cost/m^3 (MG) influent	$0.0374/$m^3$ ($142/MG)

The impacts of chemical addition on plant O&M costs are difficult to assess, since major plant modifications occurred at the time when phosphorus removal was initiated. According to plant personnel, most sludge treatment equipment operated at the plant requires approximately twice the maintenance when compared with the period prior to the beginning of phosphorus removal. Since plant flows have nearly doubled in the interim, the relative increase/MG of plant influent has been negligible.

Capital Costs--

The capital cost of the original Coldwater plant was financed by 30 year general obligation bonds providing $650,000 at 2.25 percent interest beginning October 1, 1951. The additions and modifications to the plant were financed beginning October 1, 1970 as follows:

Federal Grant	$301,500.00
State Grant	$800,000.00
Revenue Bonds	$550,000.00
Total	$1,651,500.00

The revenue bonds mature in 17 years, at an initial interest rate of 4.5 percent per annum. In 1983, the interest rate begins to increase, reaching 5.5 percent at maturity.

Summary and Conclusions

There has been a significant increase in sludge and sidestream solids concentrations entering and leaving every unit of the Coldwater, Michigan, wastewater treatment plant, beginning with the flocculation channel through the sand drying beds. The impacts of chemical addition on the treatment units were as follows:

Flocculation Channel - more solids entering in the following sidestreams:

- Secondary clarifier sludge
- Digester supernatant
- Pressure sand filter backwash water
- Ferric chloride and anionic polymer additions.

Primary Clarifiers:

- Increased average SS loading
- Increased volumes of primary sludge pumped to primary digester at higher average TS concentration
- Decreased average BOD and SS removal efficiencies.

Trickling Filters:

- Increased average BOD and TS influent and effluent concentrations
- Discoloration of filter media.

Final Clarifiers:

- Increased average SS loading
- Increased volumes of sludge pumped to flocculation channel at higher average TS concentration.

Pressure Sand Filters:

- Increased SS loading
- More backwash water at higher average SS concentration.

Digesters:

- Increased volumes of sludge feed at higher average TS concentration
- Higher supernatant average SS concentration.

Sand Drying Beds:

- Increased quantities of sludge requiring longer drying times.

In conclusion, a solids build-up has occurred throughout the plant. The additional solids have been almost entirely nonvolatile. Although O&M costs have nearly doubled since phosphorus removal began, the plant flow doubled at the same time, resulting in no increase in treatment costs per million gallons of plant influent.

CASE STUDY F: LAKEWOOD, OHIO

Introduction

The Lakewood, Ohio, experience is similar to that encountered at many older facilities which have inadequate sludge handling capacity. Soon after the plant was built in 1966, it became apparent that available flash drying capacity was inadequate, and as the equipment has aged, its capacity has further decreased. Alum addition for phosphorus removal further decreased the dryer's capacity by increasing the moisture content of the filter cake, and alum has also increased the amount of sludge to be handled. A bottleneck in the solids handling end of the plant has resulted. This had led to high solids buildup and recirculation throughout the plant and a high SS concentration in the plant effluent.

In spite of the problems created by inadequate sludge handling capacity, the plant management stayed, until 1975, with a filter and dryer operation schedule of one 8-hr shift/day because of budget considerations. It has been possible to observe the effects on plant performance of extending the operation of two shifts.

It has also been possible to observe the effects on plant performance of hauling liquid sludge from the digesters. It appears that liquid sludge hauling as a sludge management alternative is frequently relied upon by plants which have additional sludge to handle because of phosphorus removal but do not have the necessary existing sludge handling capacity.

As with other wastewater treatment plants, sludge production at Lakewood is affected by changes in wastewater flow volume. The plant was designed to treat 0.05 mil m^3 (13 MG) of wastewater per day. The actual average dry weather flow is only 0.045 mil m^3/day (12 mgd). But peak flow rates occurring during wet weather are considerably higher, since sanitary and storm sewers in the area are 40 percent combined. Table C-13 shows the average and peak influent flow rates for each year since 1973. Before September of 1975, peak flows reached almost 0.076 mil m^3/day (20 mgd). Since then, they have been kept at 0.060 mil m^3/day (16.0 mgd) by bypassing the excess to the river. Before bypassing was used, on days of heavy flow, treatment efficiency was reduced and effluent suspended solids concentrations reached 400 mg/ℓ.

TABLE C-13. PLANT INFLUENT WASTEWATER FLOW RATES, LAKEWOOD, OHIO

Year	Average		Peak	
	Total mil m^3/day	Total mgd	Total mil m^3/day	Total mgd
1973	0.055	14.6	0.075	19.8
1974	0.055	14.6	0.073	19.4
1975	0.050	13.3	0.067	17.7
1976	0.051	13.4	0.060	16.0
1977 (thru May)	0.042	11.2	0.060	16.0

History

The following description details historical plant changes which have affected sludge characteristics:

1938 - Original Imhoff plant constructed; sludge anaerobically digested and dewatered on sand drying beds.

1966 - Conventional activated sludge plant constructed; gravity thickeners, two new anaerobic digesters, two vacuum filters, and a flash dryer added to handle additional sludge.

1974 - Alum addition for phosphorus removal begun.

1975 - Sludge handling system (filters and dryer) operating schedule extended to two shifts to handle additional sludge generated.

1977 - Returned to single shift sludge handling; hauled excess liquid sludge to land disposal.

Chemical Addition for Phosphorus Removal

Liquid alum (48.4 percent $Al_2(SO_4)_3 \cdot 14\ H_2O$) is added to the mixed liquor in the activated sludge effluent channel, an open concrete channel leading to the final division well ahead of the secondary clarifiers. The alum is added through a non-submerged pipe at a point about 2 m (5 ft) from the aeration tank effluent weir and 20 m (70 ft) before the final division well.

The chemical dosage is controlled by manual adjustment of the chemical feed pump settings. The pumps are normally reset at midnight and at eight in the morning each day to compensate for the smaller wastewater flow at night. The same settings are used each day unless the plant effluent phosphorus concentration increases, in which case more chemical is added. A total of about 4,160 m^3 (1,100 gal) of alum are used each day. Each liter liquid contains 0.650 g/m^3 of $Al_2(SO_4)_3 \cdot 14\ H_2O$ (5.4 lb/gal). Therefore, the average wastewater concentration is approximately 63 ppm $Al_2(SO_4)_3 \cdot 14\ H_2O$.

Presently, the plant is not achieving the required 1 mg/ℓ effluent total phosphorus concentration because it cannot handle the additional sludge which is produced when alum dosage is increased or a polymer is used. Only 65 percent reduction of the total phosphorus concentration of the influent is achieved within the plant. The average final effluent concentration is 6.4 mg/ℓ.

General Description of Wastewater Treatment Operations

Figure C-9 shows the treatment system flow diagram. Suspended solids in the raw sewage influent plus biological and chemical solids generated by the activated sludge and phosphorus removal operations are collected by the primary and secondary clarifiers and then are gravity thickened. A portion of the solids are converted to liquid and gas by the digesters, with the remainder being either hauled to agricultural lands, dried on sand beds, or vacuum filtered and flash dried. The underdrainage from the sludge drying beds and the thickener operations are returned to the front of the plant. Other sidestreams from sludge treatment operations (digester supernatant, vacuum filter filtrate and flash dryer scrubber water) are collected and returned to the system at the primary division well.

Descriptions of sludge-significant wastewater treatment unit operations follow.

Primary Clarifiers--

The primary clarifiers consist of four circular Eimco units, with 19.8 m (65 ft) diameters, 308 m^2 (3,320 ft^2) surface areas, and 2.9 m (9.4 ft) side water depths. Each clarifier holds 883 m^3 (0.233 MG). Scum is collected in scum boxes and pumped to the primary digesters. Sludge is mechanically scraped to a hopper in the center of the sloping clarifier bottom by four revolving arms. The sludge hoppers are 0.9 m (3 ft) deep, 0.9 m (3 ft) by 0.6 m (2 ft) wide at the top, and narrowing like a square cone to a 15 cm (6 in) sludge outlet pipe at the bottom.

Although there is room in the primary clarifier bottoms, sludge storage here is avoided. Sludge is continually scraped from the bottoms of the clarifiers, and it continually flows by gravity from the hoppers to the primary sludge well. The primary clarifiers are not equipped or operated to achieve thickening, although less frequent removal of sludge could probably be practiced to allow thickening of sludge to a higher

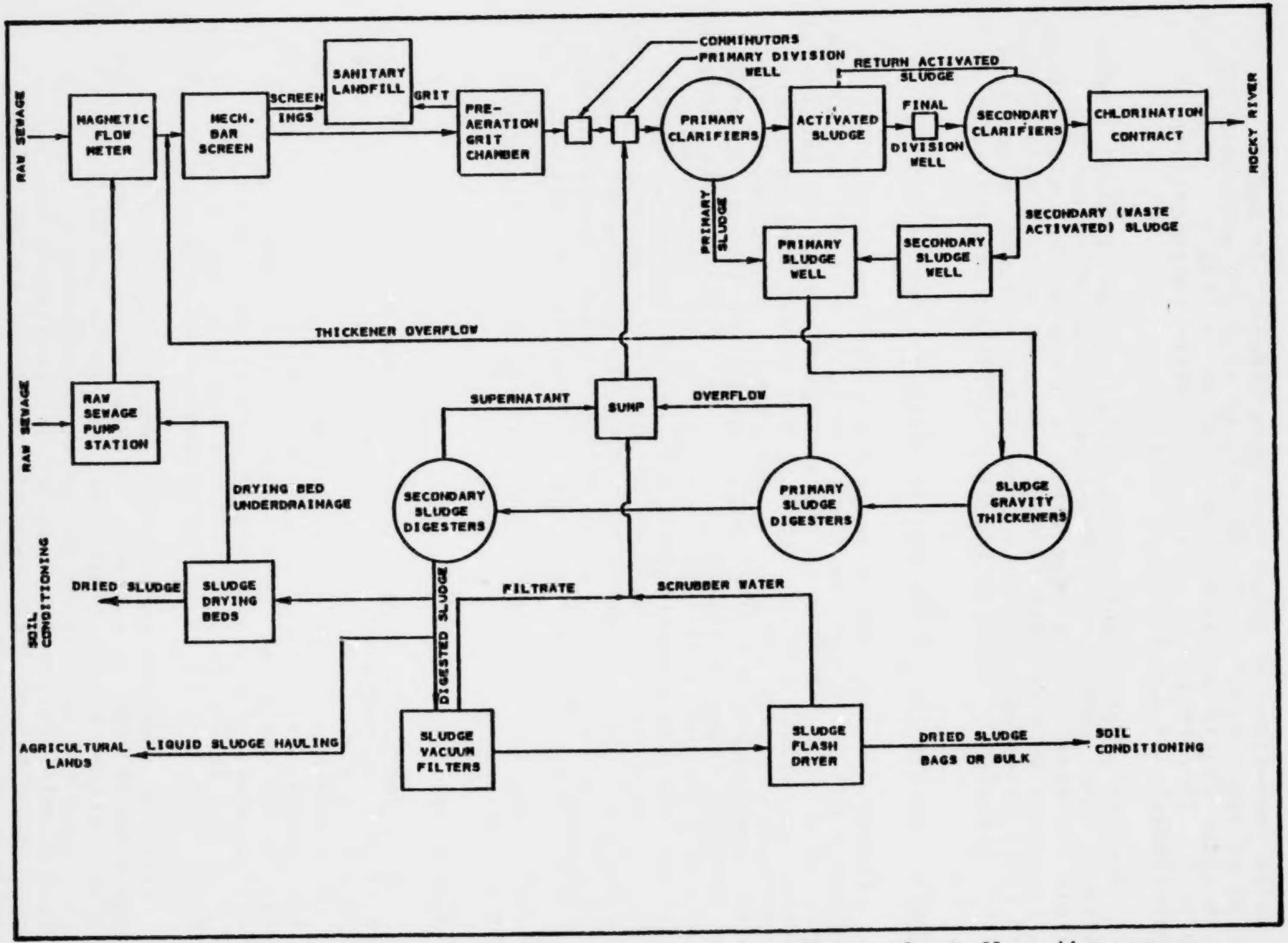

RAW SEWAGE
MAGNETIC FLOW METER
MECH. BAR SCREEN
SCREENINGS
SANITARY LANDFILL
GRIT
PRE-AERATION GRIT CHAMBER
COMMINUTORS
PRIMARY DIVISION WELL
PRIMARY CLARIFIERS
ACTIVATED SLUDGE
RETURN ACTIVATED SLUDGE
FINAL DIVISION WELL
SECONDARY CLARIFIERS
CHLORINATION CONTRACT
ROCKY RIVER
PRIMARY SLUDGE
PRIMARY SLUDGE WELL
SECONDARY SLUDGE WELL
SECONDARY (WASTE ACTIVATED) SLUDGE
THICKENER OVERFLOW
RAW SEWAGE
RAW SEWAGE PUMP STATION
SUPERNATANT
SUMP
OVERFLOW
DRYING BED UNDERDRAINAGE
SECONDARY SLUDGE DIGESTERS
PRIMARY SLUDGE DIGESTERS
SLUDGE GRAVITY THICKENERS
SOIL CONDITIONING
DRIED SLUDGE
SLUDGE DRYING BEDS
DIGESTED SLUDGE
FILTRATE
SCRUBBER WATER
AGRICULTURAL LANDS
LIQUID SLUDGE HAULING
SLUDGE VACUUM FILTERS
SLUDGE FLASH DRYER
DRIED SLUDGE
BAGS OR BULK
SOIL CONDITIONING

Figure C-9. Lakewood, Ohio, wastewater treatment plant flow diagram.

solids concentration. This is avoided, however, because the plant manager wishes to avoid problems in thickener operation which might be caused by a heavier sludge feed.

Secondary Clarifiers--

The secondary clarifiers are four circular Eimco units with 23 m (75 ft) diameters, 411 m^2 (4,420 ft^2) surface areas, and side water depths of 3.6 m (11.7 ft). Each has a 1,465 m^3 (0.387 MG) capacity. The sludge collection mechanism is the plow and scraper type like that used in the primary clarifiers. The sludge hopper is 1.2 m (4 ft) deep and 0.6 m (2 ft) by 0.9 m (3 ft) wide at the top. It then narrows to a 20 cm (8 in) sludge outlet pipe at the bottom.

The secondary clarifiers are heavily relied upon for sludge storage because the plant is overloaded with secondary sludge which has no place to go. A significant portion of this sludge is due to alum addition for phosphorus removal. A sludge blanket about 2.3 m (7.5 ft) deep has built up in the secondary clarifiers. The detention time of sludge in the secondary clarifier bottoms may be somewhere around 1.3 hr, even though sludge is continually scraped to the hoppers and continually flows by gravity to the secondary sludge well and then to the primary sludge well. Pumping of sludge from the primary and secondary sludge wells is also continuous.

Activated Sludge Aeration Basins--

The eight aeration basins are operated as conventional activated sludge treatment units with somewhat tapered aeration. Mixing and aeration both are accomplished with diffused air. Each basin holds 1,950 m^3 (0.515 MG), giving a total capacity of 15,600 m^3 (4.12 MG). Primary effluent and return activated sludge enter and flow the lengths of four of the eight basins, then flow down the other four basins before emptying into the effluent channel.

Because the gravity sludge thickener is overloaded, 97 to 98 percent of the secondary sludge which is produced is recirculated to the aeration basins as return activated sludge. This has resulted in a high mixed liquor suspended solids concentration.

Detailed Description of Sludge Treatment and Disposal Operations

Thickening--

The two circular Eimco gravity thickeners have 10.7 m (35 ft) diameters and 3 m (10 ft) side water depths. The thickeners were designed to operate with an influent total solids concentration of less than 0.2 percent. Mechanisms for sludge stirring or polymer addition are absent. Sludge is fed at the center of the top of each clarifier. Two collector arms on the bottom of each thickener scrape the thickened sludge toward circular center hoppers which are approximately 1.1 m (36 ft 6 in) in diameter and 0.6 m (2 ft) deep. Sludge is pumped from the hoppers to the digesters. Sludge from the #1 thickener is pumped to the #1 primary digester, while that from the #2 thickener is pumped to the #2 primary digester.

The sludge thickener is continuously fed combined primary and secondary from the primary sludge well at a rate of 3.240 m^3/min (850 gal/min). Sludge is removed from the thickener intermittently at 0.282 m^3/min (75 gal/min) for 15 min each hour. The top of the sludge layer in each thickener is near the overflow weir. This overloaded condition is responsible for solids carryover and frequent thickener bulking.

Digestion--

There are four above-ground anaerobic digestion tanks. The primary digesters are 22.9 m (75 ft) in diameter with a 6.8 m (22.5 ft) side water depth, while the secondary digesters are 18.3 m (60 ft) in diameter with a side water depth of 5.8 m (19 ft). The sludge flow pattern through the digesters is shown in Figure C-10. Sludge flows by gravity from the primaries to the secondaries.

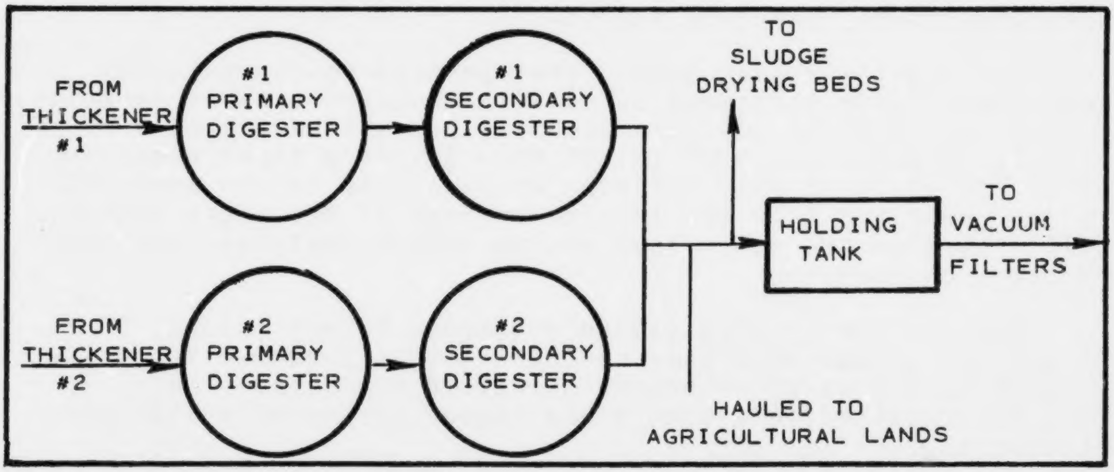

FROM THICKENER #1
#1 PRIMARY DIGESTER
#1 SECONDARY DIGESTER
TO SLUDGE DRYING BEDS
HOLDING TANK
TO VACUUM FILTERS
FROM THICKENER #2
#2 PRIMARY DIGESTER
#2 SECONDARY DIGESTER
HAULED TO AGRICULTURAL LANDS

Figure C-10. Lakewood, Ohio, anaerobic digester configuration.

The primary digesters are mixed and heated. The mixing mechanism is a "circulating mixer" consisting of a propeller on a shaft inside of a draft tube. Each primary digester has three circulating mixers, each capable of circulating at least 473 ℓ/sec (7,500 gal/min) of water.

The temperature in the primary digesters is maintained at 77.2 to 87.2°C (95 to 105°F). Heating is accomplished by recirculating sludge through two external heat exchangers. Sludge recirculation through the heat exchangers also provides limited mixing of the contents of the primary digesters. The two hot water boilers serving the heat exchangers can run on either methane digester gas or No. 2 fuel oil. Throughout most of the year, digester gas alone is used. It is consumed in the winter at a rate of 896 to 1,060 m^3/day (32,000 to 38,000 ft^3/day) per boiler and at a rate of 360 to 560 m^3/day (13,000 to 20,000 ft^3/day) in the summer. In addition, approximately 3.785 m^3/yr (1,000 gal/yr) fuel oil are needed for digester heating.

The primary digesters have uninsulated anchored steel dome covers while the secondaries have insulated floating covers. Methane gas is collected from all four digesters and compressed into two storage spheres for use in digester and building heating.

The secondary digesters are neither mixed nor heated. Sludge in the secondaries stays warm at about 87°F due to previous heating in the primaries. Supernatant can be drawn from the secondary digesters at three different levels, 1.8 m (6 ft), 1.5 m (5 ft), and 1.2 m (4 ft) from the top.

Each primary digester is fed from one thickener for 15 min/hr, 24 hr/day. Therefore each primary digester is fed for a total of 6 hr/day at 0.284 m^3/min (75 gal/min), 7 days/wk. Sludge overflow from the primaries to the secondaries and supernatant overflow from the secondaries is more or less continuous by gravity. From each secondary digester, 19 m^3 (5,000 gal) of sludge are pulled to tank trucks 4 times a day, 5 days a week. In addition, sludge flows by gravity 0.144 m^3/min (38 gal/min) from each secondary digester to the holding tank for approximately 3 hr and 15 min/day, 5 days/wk. When the filter and dryer was operated 2 shifts a day, as before liquid sludge hauling was practiced, the flow rate from each digester was about 0.16 m^3/min (42 gal/min) about 7 hr/day, 5 days/wk. Finally, each sand drying bed is filled twice a year with digested sludge.

The digested sludge holding tank is a simple 49,000-m^3 (13,000-gal) tank. It is unmixed, and no dilution water is added to it or supernatant removed.

The digesters have not been cleaned since they were installed in 1966, and their effective volume is presumed to be considerably reduced from their total volume. There is evidence that the sides and corners of the secondary digesters are filled in with a very heavy sludge (estimated at about 15 percent solids).

Dewatering-Vacuum Filters--

Two Eimco belt-type rotary drum vacuum filters are used for dewatering digested sludge. Each filter has a drum diameter of 3 m (10 ft), a length of 3.6 m (12 ft), and filtering area of 35 m^2 (376 ft^2). The filters have 24 drum compartments. The vacuum pumps have a capacity of 15 m^3/min (545 ft^3/min) at 51 cm (20 in) Hg vacuum. The filter drive motors are variable-speed, enabling a drum speed of 1.5 to 9 min per revolution. The filter medium is polyethylene cloth. The filter tanks are equipped with swing-type agitators capable of completing 6 to 18 cycles/min.

The design filtering performance specifies a capability of filtering a sludge of 1.5 to 3 percent solids at a rate of 35 to 65 kg/m^2/hr (1.25 to 2.25 lb/ft^2/hr) when properly conditioned with ferric chloride and lime. The maximum suspended solids concentration of the filtrate is specified as 500 ppm.

Only one vacuum filter can be operated at a time because of the limiting capacity of the flash dryer. One filter operates 5 days/wk and for 6.5 hr/day when operated one shift, and 13 hr/day when operated two shifts. Conditioning chemical dosages are roughly 137 kg dry lime/t of dry solids (275 lb dry lime/ton), and 0.283 m^3 (63 gal) liquid ferric chloride (40 percent $FeCl_3$) or 15 kg $FeCl_3$/t of dry solids (30 lb $FeCl_3$/t). The ferric chloride is diluted to 17 percent before mixing with sludge in the flash mixer, and the lime is slaked. The lime is purchased as pebble lime, 72 percent CaO (calcium oxide). The filter drum speed has been set at 8 min/revolution and the depth of drum submergence is around 0.76 to 0.91 m (30 to 36 in).

Dewatering-Drying Beds--

The six sludge drying beds are greenhouse-enclosed sand beds which are ventilated through open doors and windows. The beds consist of 23 cm (9 in) of sand underlain by 8 to 23 cm (3 to 9 in) of gravel. The effective size of the filter sand is 0.25 to 0.50 mm with a uniformity coefficient not greater than 4.0. Bed filtrate collects underneath in a lengthwise row of drain tile under the middle of each bed. These conduits empty into a long intersecting drain which runs to the raw sewage lift station. The beds are 8.5 m (27 ft 8 in) wide, and 50 m (163 ft) long. The concrete bed walls rise 0.5 m (1.5 ft) above the sand. Dried sludge is lifted from the beds manually with a fork. The dried sludge is then made available to the public. Interested persons can drive into the plant and pick up the sludge, or it can be delivered to them in a 2-1/2-ton dump truck.

Sludge is applied to the drying beds twice a year. The first application is in the spring and the second at the end of the summer. All but two beds are filled at these times. The two beds are available for use later, should the filter and dryer system be down for repair.

Heat Drying--

The plant has a flash drying system by C-E Raymond Combustion Engineers. Filter cake is transported by a belt conveyor to the dryer. The design capacity of the dryer is 825 kg dry solids/hr (1,820 lb/hr). The design is based upon the following sludge cake feed characteristics:

- Moisture content 75 percent
- Volatile matter 51.4 percent
- 3,000 kg cal/kg (5,400 Btu/lb) of dry solids.

The furnace burns either No. 2 fuel oil or methane digester gas.

Sludge drops off the conveyor into a pug mill mixer where it is thoroughly mixed with dried sludge. The sludge then passes through a cage mill where it is pulverized and entrained with rising air and gases which have been heated in a furnace. Sludge and hot gases rise 9 m (30 ft) through a vertical duct to a cyclone separator. The dried sewage solids are separated from the hot gases and dust which are drawn off by an induced draft fan. The hot gases go either back down to the pug mill to be recycled through the system or they go to a deodorizing preheater where some of their heat is recovered in preheating combustion air. From the preheater, the gases go into the furnace to be deodorized. Then they pass through the combustion air preheater for further heat recovery before entering the wet scrubber.

The wet scrubber consists of a wet centrifugal precleaner followed by a dynamic precipitator with a water spray. The dust-laden scrubber water is discharged and returned to the head of the treatment plant.

The dried sludge passes from the cyclone separator through an airlock, and 80 to 85 percent of it is recycled back to the pug mill to be mixed with wet filter cake while 15 to 20 percent is conveyed to storage.

The dryer is furnished with a pneumatic dried sludge cooling and conveying system to receive the product from the drying system and deliver it to the storage bin. It consists of a cyclone collector, fan, and wet scrubber. There is also a dried sludge bagging machine and bulk loading conveyor. Interested persons can pick up the bagged product at the plant for use on lawns, gardens, etc. Deliveries are made by the plant to bulk users in a truck that carries 4 tons.

The flash dryer operates whenever sludge is being filtered. However, the furnace is started up an hour before filtering begins. When sludge is being fed, the furnace is at a temperature of roughly 815°C (1,500°F), and the hot gases are at about 593°C (1,100°F) when they are mixed with the sludge feed. The furnace is run on No. 2 fuel oil at 210,900 kg/m^2 (300 psi) pressure or occasionally on methane gas at 4,218 kg/m^2 (6 psi) pressure. When drying two shifts per day, approximately 3.785 m^3/day (1,000 gal/day) fuel oil are consumed; when drying one shift per day, about 2.08 m^3/day (550 gal/day) are consumed.

Disposal of Liquid Sludge--

The plant has contracted a private firm to haul liquid digested sludge. The company specializes in hauling liquid municipal sludge, lime water sludge, and septic tank sludge. Hauling of Lakewood liquid sludge has been going on only since June of 1977. A "tee" was installed on the line running from the digesters to the holding tank, so that sludge could be sucked directly into a tank truck with a vacuum pump. When the sludge is heavy, a piston pump, normally used to pump to the drying beds, is also applied. The tank trucks which are used hold 17,000, 18,900, or 20,800 m^3 (4,500, 5,000, or 5,500 gal) of sludge.

Sludge is hauled approximately 115 km (70 mi) to agricultural lands outside of the Cleveland suburbs. From the tank truck the sludge is forced into a special truck or "field jimmy" which is adapted for spreading liquid sludge on fields.

Normally eight 19,000-m^3 (5,000-gal) loads of digested sludge are hauled per day, 5 days/wk.

Impact of Phosphorus Removal on Wastewater Treatment Operations

Alum is added for phosphorus removal to the channel between the aeration basins and the final clarifiers. Final clarification is directly affected by the alum by increasing the capture of SS and BOD. The resulting chemical sludge indirectly affects primary and secondary clarification and activated sludge treatment by altering the composition of the sidestreams and flows entering these processes.

Wastewater samples for SS, BOD, and phosphorus analyses are taken at an insufficient number of points in the treatment process to allow a complete analysis of the impact of the chemical sludge on wastewater treatment operations. For example, a single sampling point exists before the primary clarifiers, ahead of the point where the sidestreams from the sludge treatment processes join the raw sewage. There is no sample of the wastewater which actually enters the primary clarifiers.

Table C-14 presents data taken from plant monthly reports on wastewater SS, BOD, and total phosphorus and orthophosphorus concentrations. Routine samples for SS and BOD determination are taken of the raw sewage at the magnetic flow meter building, of the primary clarifier effluent, and of the final effluent. Phosphorus determinations are made only on the raw sewage and final effluent. The table presents the average concentrations for 1 yr with no alum addition, and 1 yr during which 63 mg/ℓ was added to the sewage.

There is only a small difference between the plant influent and primary effluent SS and BOD figures. This indicates that the return sidestreams from the thickeners and digesters are contributing heavy solids loadings to the primary clarifiers. The solids accumulate in the primary clarifiers from whence they are fed back to the thickeners and digesters, thus being trapped essentially in a closed-loop system. Table C-14 shows that the primary effluent SS concentration was higher when alum was added than before, presumably due to the heavier solids loading contributed by the sidestreams entering the primary clarifiers.

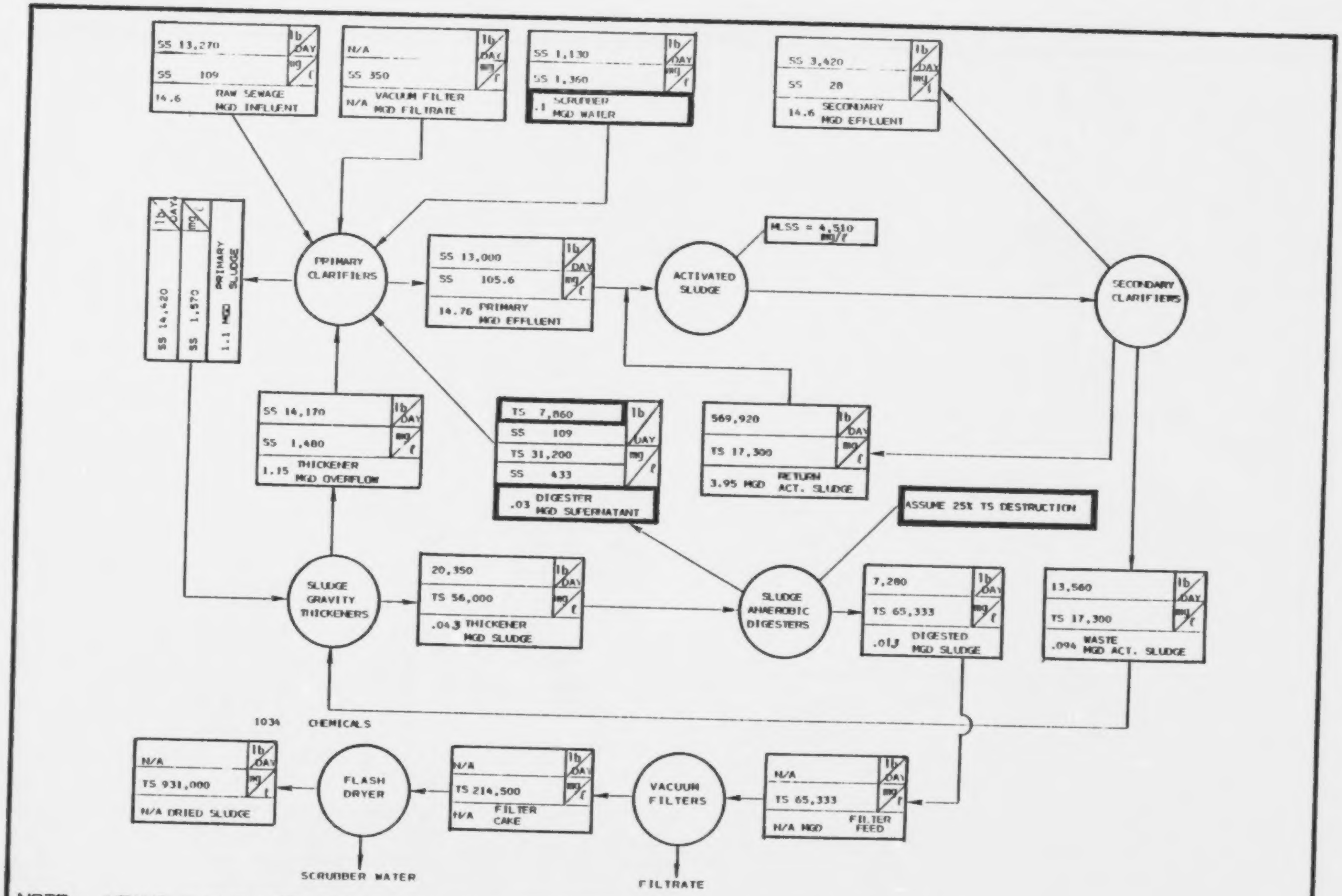
Suspended solids mass balances were constructed about the primary clarifiers to the extent possible with available plant data. Figure C-11 is a mass balance which is derived from a year of performance data during which 63 mg/ℓ alum addition was practiced. Figure C-12 portrays the situation for a year with no alum addition, but data for this period is incomplete.

The information in Figure C-11 shows that the sidestreams are contributing a loading of 6,948 kg/day (15,305 lb/day) SS to the primary clarifiers. This figure is greater than the 6,025 kg/day (13,272 lb/day) contained in the plant influent. By far the most significant sidestream in terms of solids input is the thickener overflow, carrying 6,433 kg/day (14,171 lb/day) SS. The plant superintendent has indicated that the thickener overflow quality deteriorated when alum addition was started, but the average increase in solids concentration is unknown. The digester supernatant also contributes a large amount of solids, and a comparison of Figures C-11 and C-12 shows that the supernatant quality was poorer during alum addition.

During alum addition, higher SS and BOD loadings were carried into the aeration basins in the primary effluent. Table C-15 shows that the plant's final effluent concentrations did not change significantly, however. This indicates that the addition of alum allowed increased capture of solids and BOD in the final clarifiers. Table C-15 shows the "removals" (or the differences between primary and final effluent concentrations) of SS and BOD during secondary treatment with and without alum addition. There were on the average 0.015 kg/m^3 (125 lb/MG) additional SS and 600 kg/mil m^3 (5 lb/MG) additional BOD removed during alum addition. These figures do not give the whole

TABLE C-14. PLANT SS, BOD, AND PHOSPHORUS CONCENTRATIONS BEFORE AND DURING ALUM ADDITION FOR PHOSPHORUS REMOVAL, LAKEWOOD, OHIO

	No alum addition (Nov. 72 to Oct. 73)	63 mg/ℓ alum addition (Jan. 74 to Dec. 75)
Wastewater flow m^3/day (mgd):	55,700 (14.9)	54,600 (14.6)
Plant influent SS (mg/ℓ):	111	109
Primary effluent SS (mg/ℓ):	94.6	105.6
Final effluent SS (mg/ℓ):	32	28
Plant effluent BOD_5 (mg/ℓ):	97	108
Primary effluent BOD_5 (mg/ℓ):	84	84.6
Final effluent BOD_5 (mg/ℓ):	13	13
Plant influent total phosphorus (mg/ℓ):	11.8	18.29
Effluent total phosphorus (mg/ℓ):	9.5	6.4
Plant influent ortho-phosphorus:	6.9	8.6
Effluent ortho-phosphorus:	8.2	4.15


SS 13,270
SS 109
14.6 RAW SEWAGE MGD INFLUENT
N/A
SS 350
N/A VACUUM FILTER MGD FILTRATE
SS 1,130
SS 1,360
.1 SCRUBBER MGD WATER
SS 3,420
SS 28
14.6 SECONDARY MGD EFFLUENT
lb/DAY
mg/ℓ
SS 14,420
SS 1,570
1.1 MGD PRIMARY SLUDGE
PRIMARY CLARIFIERS
SS 13,000
SS 105.6
14.76 PRIMARY MGD EFFLUENT
ACTIVATED SLUDGE
MLSS = 4,510 mg/ℓ
SECONDARY CLARIFIERS
SS 14,170
SS 1,480
1.15 THICKENER MGD OVERFLOW
TS 7,860
SS 109
TS 31,200
SS 433
.03 DIGESTER MGD SUPERNATANT
569,920
TS 17,300
3.95 MGD RETURN ACT. SLUDGE
ASSUME 25% TS DESTRUCTION
SLUDGE GRAVITY THICKENERS
20,350
TS 56,000
.043 THICKENER MGD SLUDGE
SLUDGE ANAEROBIC DIGESTERS
7,280
TS 65,333
.013 DIGESTED MGD SLUDGE
13,560
TS 17,300
.094 WASTE MGD ACT. SLUDGE
1034 CHEMICALS
N/A
TS 931,000
N/A DRIED SLUDGE
FLASH DRYER
SCRUBBER WATER
N/A
TS 214,500
N/A FILTER CAKE
VACUUM FILTERS
FILTRATE
N/A
TS 65,333
N/A MGD FILTER FEED

NOTE: HEAVILY OUTLINED BOXES CONTAIN ESTIMATED VALUES. N/A = NOT AVAILABLE.

Figure C-11. Lakewood, Ohio, hydraulic and solid mass balance during 63 mg/ℓ alum addition (Jan through Dec 74)

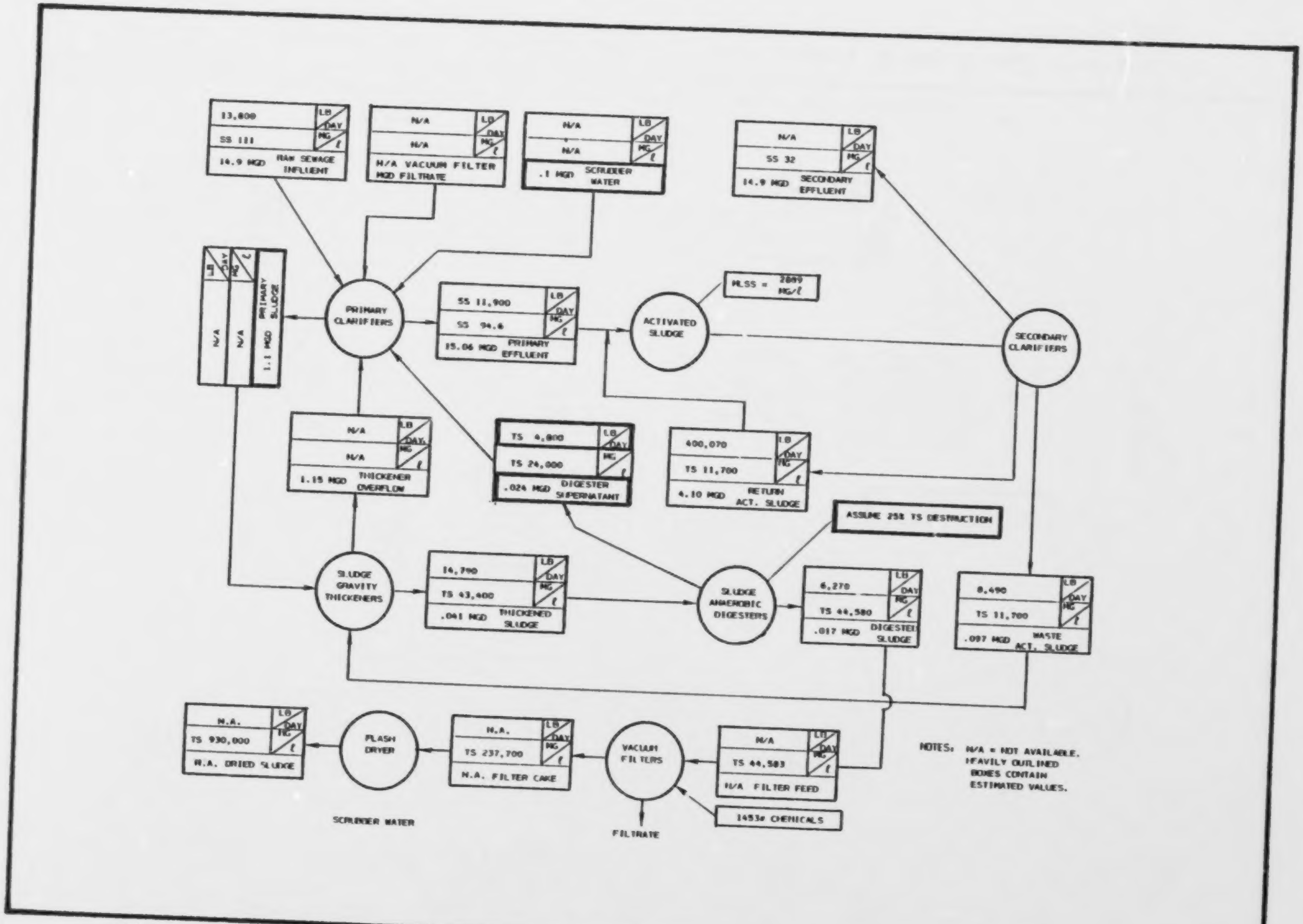

13,800
SS 111
14.9 MGD RAW SEWAGE INFLUENT
N/A
N/A
N/A VACUUM FILTER MGD FILTRATE
N/A
N/A
.1 MGD SCRUBBER WATER
N/A
SS 32
14.9 MGD SECONDARY EFFLUENT
LB DAY
MG ℓ
N/A
N/A
1.1 MGD PRIMARY SLUDGE
PRIMARY CLARIFIERS
SS 11,900
SS 94.6
15.06 MGD PRIMARY EFFLUENT
ACTIVATED SLUDGE
MLSS = 2849 MG/ℓ
SECONDARY CLARIFIERS
N/A
N/A
1.15 MGD THICKENER OVERFLOW
TS 4,800
TS 24,000
.024 MGD DIGESTER SUPERNATANT
400,070
TS 11,700
4.10 MGD RETURN ACT. SLUDGE
ASSUME 25% TS DESTRUCTION
SLUDGE GRAVITY THICKENERS
14,790
TS 43,400
.041 MGD THICKENED SLUDGE
SLUDGE ANAEROBIC DIGESTERS
6,270
TS 44,580
.017 MGD DIGESTED SLUDGE
8,490
TS 11,700
.097 MGD WASTE ACT. SLUDGE
N.A.
TS 930,000
N.A. DRIED SLUDGE
FLASH DRYER
N.A.
TS 237,700
N.A. FILTER CAKE
VACUUM FILTERS
N/A
TS 44,583
N/A FILTER FEED
1453# CHEMICALS
FILTRATE
SCRUBBER WATER
NOTES: N/A = NOT AVAILABLE. HEAVILY OUTLINED BOXES CONTAIN ESTIMATED VALUES.

Figure C-12. Lakewood, Ohio, hydraulic and solid mass balance before alum addition (Nov 72 through Oct 73)

TABLE C-15. PLANT SS, BOD, AND PHOSPHORUS REMOVAL BEFORE AND DURING ALUM ADDITION FOR PHOSPHORUS REMOVAL, LAKEWOOD, OHIO

	Secondary Treatment		
Removal	No Alum	63 mg/ℓ Alum	Difference
kg SS/m^3 (lb/MG)	0.06 (522.1)	0.077 (647.2)	0.017 (125.1)
kg BOD/m^3 (lb/MG)	0.071 (592.1)	0.072 (597.1)	0.001 (5.0)
% SS	66.2	73.5	7.3
% BOD	84.5	84.6	0.1

	Total Plant		
Removal	No Alum	63 mg/ℓ Alum	Difference
kg P/m^3 (lb/MG)			
Total	0.002 (19.2)	0.018 (15.2)	0.016 (133.3)
Ortho	0.0013 (10.8)	0.004 (37.1)	0.006 (47.9)
% P			
Total	19.5	65.0	45.5
Ortho	18.8	51.7	70.5
kg SS/m^3 (lb/MG)	0.08 (658.9)	0.081 (675.5)	0.002 (16.6)
kg BOD/m^3 (lb/MG)	0.084 (700.5)	0.095 (792.3)	0.010 (91.8)
% SS	71.2	74.3	3.1
% BOD	86.6	88.0	1.4

picture, however, because the aeration basin influent includes the return activated sludge in addition to the primary effluent. The mass balance figures, C-11 and C-12, indicate the role of the return activated sludge solids. Because of a higher TS concentration in the return activated sludge, and an increase in the mass of solids returned during alum addition, the solids loading to the aeration basins went up and the mixed liquor suspended solids concentration rose by 56 percent. With this information, it is possible to calculate the SS mass removed in the final clarifiers before and after alum addition. The MLSS concentration is taken as the SS concentration of the final clarifier influent. Effluent SS concentration and wastewater flow rate are also known. The mass of SS removed can thus be calculated as 2.85 kg/m^3 (23,827 lb SS/MG) before alum addition, and 4.48 kg/m^3 (37,380 lb SS/MG) during alum addition. The difference is 1.63 kg/m^3 (13,553 lb SS/MG). On a per-day basis, there were 86.1 kkg/day (95 tons/day) more secondary sludge SS generated during alum addition.

Impact of Phosphorus Removal on Sludge Treatment and Disposal Operations

Additional Sludge Quantity Produced by Alum Addition--

Figures C-11 and C-12 summarize all available information on volume and mass changes in sludges and sidestream quantities after alum addition. Data contained in monthly reports indicate that as a result of alum addition to the channel between the aeration basins and the secondary clarifiers, the average TS concentration of the secondary sludge rose from 1.17 to 1.73 percent TS. The volume of secondary sludge remained the same, however. After alum addition, approximately 0.015 mil m^3/day (4.1 mgd) secondary sludge were still generated, with 0.015 mil m^3/day (4 mgd) still being returned to the aeration basins and 0.004 mil m^3/day (0.1 mgd) still being wasted to the gravity thickeners. Because of the higher solids concentration of the sludge, however, calculation shows that an average of 78.9 kkg/day (87 ton/day) more secondary sludge TS were generated, or 1.5 kg/m^3 (12,554 lb/MG). This is similar to the estimate of 86.1 kkg/day (95 ton/day), or 1.62 kg/m^3 (13,553 lb/MG) additional sludge SS arrived at in the last section by a calculation of mass balance.

Of the 78.9 kkg/day (87 ton/day) additional sludge TS, approximately 76.6 kkg/day (84.5 ton/day) are returned to the aeration basins, while approximately 2.26 kkg/day (2.5 ton/day) are wasted to the thickeners.

Operational Problems Attributable to the Production of Chemical Sludge--

Gravity thickeners--Alum addition created essentially no change in the hydraulic balance of the thickener. However, the solids concentrations and, therefore, the solids loads of the sludge feed, the thickener overflow, and the thickened sludge, were increased.

Approximately 2,270 kg/day (5,000 lb/day) of additional secondary sludge TS were pumped to the gravity thickeners after alum addition. Figures C-11 and C-12 show that the mass of gravity thickened sludge which was pumped to the digesters increased by a similar amount. The total solids concentration of the thickened sludge which was pumped to the digesters rose from 4.34 to 5.6 percent.

It is likely that the average SS concentration of the thickener overflow was higher after alum addition than before, due to more severe thickener overloading and more frequent thickener bulking. The thickener overflow, heavily loaded with solids, is returned to the primary clarifiers where it contributes to primary sludge. Since the primary sludge is pumped back to the thickeners, a circular flow path exists. A heavy solids load is recirculated along this path.

The thickeners were designed for a surface settling rate of 33 m^3/day/m^2 (811 gal/day/ft^2). Both before and after alum addition, the actual rate was much higher than this, about 46.6 m^3/day/m^2 (1,143 gal/day/ft^2). According to the plant superintendent, the thickener was designed for a sludge feed of less than 2,000 ppm TS. The actual feed concentration was at least 2,400 ppm before and during alum addition. A sludge blanket ranging from about 1.8 m (6 ft) to the entire water depth of the thickener has built up. The supernatant SS concentration ranged from 22 to 3,974 ppm in the course of 1 mo during alum addition. Thickener bulking occurs frequently with periods of stable operation in between. Bulking may have been somewhat less frequent before alum addition according to the plant superintendent.

Digesters--The volume of sludge pumped to the digesters was essentially unchanged by alum addition. The hydraulic balance of the digesters was altered, however, because the volume of removed sludge decreased while the volume of supernatant increased. Presumably, the volume of sludge removed was decreased because of reduced filter and dryer capacity.

The TS concentration of the supernatant increased to the point that it was almost as high in solids as the digesting sludge itself. This occurred because there was little room in the digesters for solids-liquid separation.

The average volatile solids concentration as a percentage of total solids of the sludge fed to the primary digesters decreased from 63.8 before alum addition to 57.8 with alum addition. The percent volatile solids of the digested sludge decreased from 50.2 before alum addition to 44.8 with alum addition. The percent volatile reduction, as calculated by the formula

$$VR = 100 \times \left(1 - \frac{\%\ \text{ASH in raw} \times \%\ \text{VS in digested}}{\%\ \text{VS in raw} \times \%\ \text{ASH in digested}}\right)$$

as 42.8 before alum addition and 40.7 percent with alum addition. The average gas production during one year of operation before phosphorus removal was 1,994 m^3/day (71,200 ft^3/day), while the average production of methane during 1 yr with alum addition was 2,178 m^3/day (77,800 ft^3/day). The digester gas characteristics were not measured before alum addition, but a single recent analysis is available showing the following composition:

Carbon dioxide (CO_2)	33%
Methane (CH_4)	67%
Carbon monoxide	0.01%
Hydrogen sulfide (H_2S)	100 ppm
Sulfur dioxide	2 ppm
Nitrogen dioxide (NO_2)	<0.1

A recent study analysis also showed the following composition of the digested sludge:

Potassium (k)	0.08%
Total Kjeldahl nitrogen	646 µg/gram dry basis
Nitrate nitrogen	23.75 µg/gram dry basis
Total phosphate	1,064 µg/gram
Cadmium	0.32 ppm
Zinc	1,475 ppm
Copper	25.3 ppm
Chromium	13.5 ppm
Mercury	0.135 ppm
Lead	1.25 ppm

Vacuum filtering and flash drying--Unfortunately, reliable measurements of flow rates or volumes of filter feed, filtrate, filter cake, and dried sludge are unavailable for before 1977. Total solids concentrations of filter feed and filter cake are available, however. The average TS concentration of the digested sludge (filter feed) rose from 4.45 percent when no alum was used to 6.53 percent during alum addition. Despite this increase in feed solids concentration, the filter cake which was produced during alum addition was wetter than it had been before. Average cake TS dropped from 23.8 percent to 21.4 percent.

Since measurements of the volume or mass of sludge filtered are unreliable for years prior to 1977, the impact of alum addition on chemical conditioner dosages required for vacuum filtering cannot be determined. During the first 5 mo of 1977, the average lime dosage used was 214 kg/kkg (429 lb/ton) of dry solids and the average ferric chloride dosage was 25.5 kg/kkg (51 lb/ton) of dry solids.

Average filter yield during this period was 0.94 kg (2.08 lb sludge solids/ft^2/hr). The filter yield can be assumed to have decreased as a result of alum addition, because filter yield at the plant cannot exceed the rate at which cake can be fed to the dryer. The rate at which cake could be dried necessarily decreased with alum addition, because the sludge cake was wetter.

Problem resolution--The basic problem facing the plant was its inability to move sludge all the way through and out of the system at a fast enough rate. Operation of the vacuum filters and dryer for one shift each day and use of the drying beds permitted the removal of only about 49.2 to 64.3 m^3/day (13,000 to 17,000 gal/day) digested sludge from the digesters. By running the filters and dryer 14 hr/day, instead of 6.5, it became possible to process 98.4 to 128.7 m^3/day (26,000 to 34,000 gal/day) digested sludge. The most obvious plant improvement which resulted was the reduction of the volume of supernatant. Pumping less supernatant reduced the load of solids circulating and recirculating through the primary clarifiers, thickeners, and digesters as sludge, thickener overflow, and supernatant. An improvement in plant effluent quality resulted. The average SS concentration of the final effluent quality resulted. The average SS concentration of the final effluent was only 18 mg/ℓ in 1976 when operating for two shifts/day.

The year 1976 was a period of operation during which the filter and dryer were run on two shifts. Table C-16 presents average values for various items recorded by the plant during this period. Based on this information, Figure C-13 presents a solids mass balance which was constructed around the digesters. Comparing it with the mass balances in Figures C-11 and C-12 indicates that removal of sludge solids and sludge gallons from the digesters increased when the plant went to two shifts, while the quantity of sludge feed stayed about the same. This led to fewer gallons of supernatant returned to the primaries. It also resulted in more room in the digester for solids-liquid separation. A higher quality supernatant resulted, with only about 1.5 percent solids rather than the previous 3.1 percent.

In conclusion, by extending the time of filter-dryer operation, the adverse impacts of alum addition on digester supernatant were overcome. Poor supernatant quality probably had been one of the factors responsible for thickener bulking according to the plant superintendent, so the adverse effect of alum addition on the thickeners was probably also overcome by running two shifts. However, even with these improvements, the plant was still overloaded with sludge, as it had been even before alum addition, and some of the adverse effects of alum addition had not been overcome.

As a solution, the hauling of liquid digested sludge was begun, and filter-dryer operation was returned to one shift per day. Liquid sludge hauling began at the beginning of June of

TABLE C-16. 1976 SLUDGE TREATMENT DATA DURING DOUBLE SHIFT VACUUM FILTER AND FLASH DRYER OPERATION AT LAKEWOOD, OHIO

Average sewage flow m^3/day (gal/day)	0.05	(13.4)
Sludge fed to digester m^3/day (gal/day)	187	(49,424)
TS digester feed (%)		4.6
Digested sludge removed m^3/day (gal/day)	85.3	(22,530)
TS digested sludge (%)		5.8
Sludge filtered m^3/day (gal/day)	84.7	(22,393)
Hours of filter operation	12.5 hr/day, 5 day/wk	

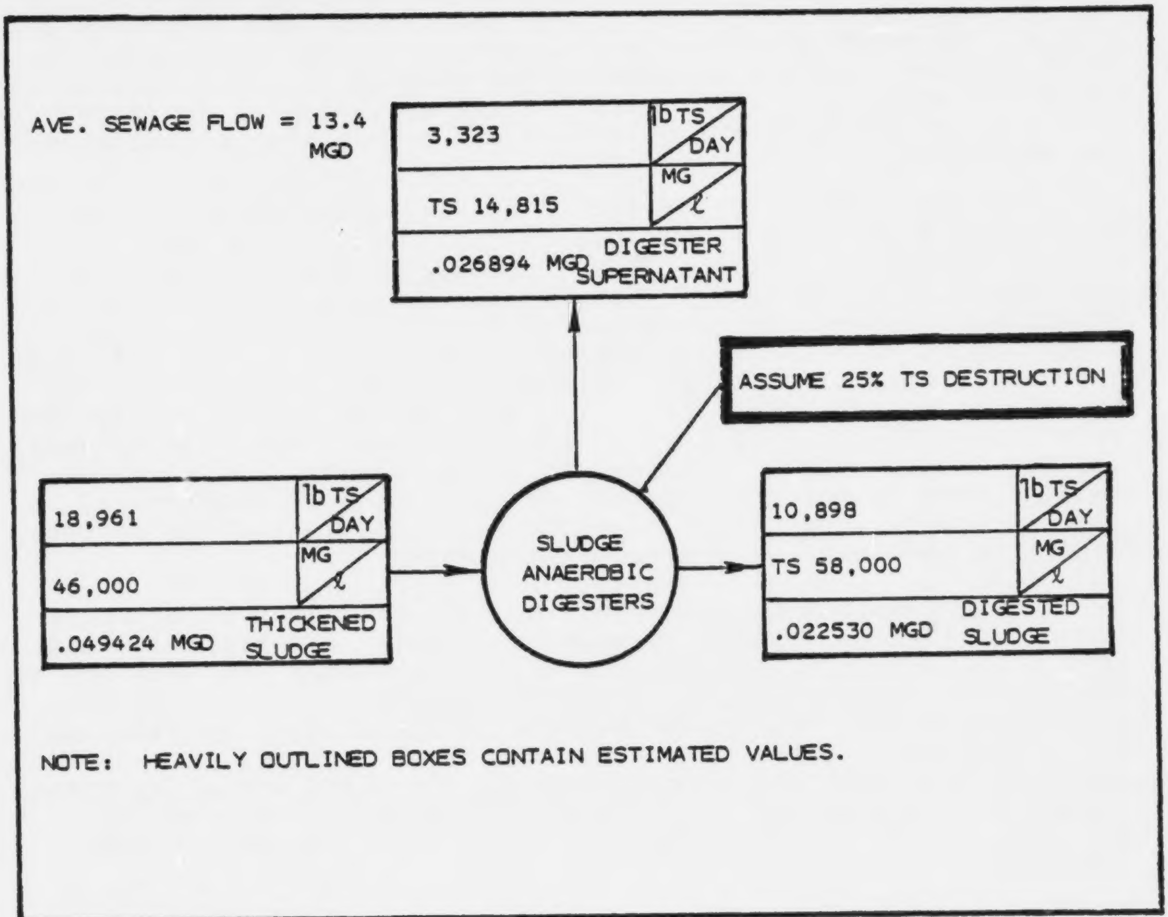

AVE. SEWAGE FLOW = 13.4 MGD
3,323
lb TS/DAY
TS 14,815
MG/ℓ
.026894 MGD
DIGESTER SUPERNATANT
ASSUME 25% TS DESTRUCTION
18,961
lb TS/DAY
46,000
MG/ℓ
.049424 MGD
THICKENED SLUDGE
SLUDGE ANAEROBIC DIGESTERS
10,898
lb TS/DAY
TS 58,000
MG/ℓ
.022530 MGD
DIGESTED SLUDGE
NOTE: HEAVILY OUTLINED BOXES CONTAIN ESTIMATED VALUES.

Figure C-13. 1976 anaerobic digestion mass balance during double shift vacuum filter and flash dryer operation at Lakewood, Ohio.

1977, so limited data are available on its effects. Figure C-14 shows that an average of 148 m^3 sludge/day (39,096 gal sludge/day) were removed from the digester in June 1977. The average quantity of sludge filtered/day was 32.3 m^3 (8,546 gal). The remaining volume of digested sludge, 115.6 m^3/day (30,550 gal/day), was hauled out as liquid. While the quantity of sludge pumped into the primary digesters was not increased by the plant operators, more sludge was removed than ever before. In fact, the sludge removal rate exceeded the sludge feed rate.

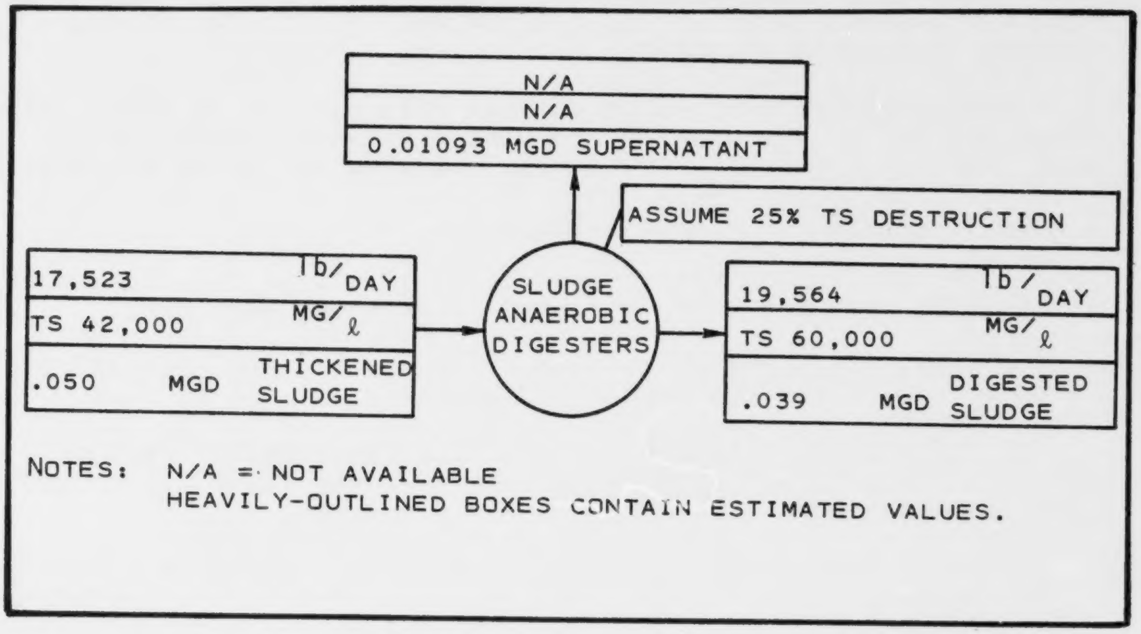

N/A
N/A
0.01093 MGD SUPERNATANT
ASSUME 25% TS DESTRUCTION
17,523 lb/DAY
TS 42,000 MG/ℓ
.050 MGD THICKENED SLUDGE
SLUDGE ANAEROBIC DIGESTERS
19,564 lb/DAY
TS 60,000 MG/ℓ
.039 MGD DIGESTED SLUDGE
NOTES: N/A = NOT AVAILABLE
HEAVILY-OUTLINED BOXES CONTAIN ESTIMATED VALUES.

Figure C-14. Anaerobic digestion mass balance during liquid sludge hauling, Lakewood, Ohio.

Removal of sludge from the digesters at this rate resulted in less supernatant return and better solids-liquid separation in the secondary digesters. The quantity of supernatant decreased to 41.4 m^3/day (10,930 gal/day), and if the TS concentration is assumed to have remained at 1.5 percent, then only 620.6 kg TS/day (1,367 lb TS/day) were returned to the primaries in the supernatant. The impacts on plant operation included a reduction of the mass of primary sludge. Consequently, more secondary sludge could be wasted to the thickeners. This was done, with the result that the MLSS concentration in the aeration basins came down to about 3,500 to 4,000 ppm, compared to 4,500 ppm when the plant was removing less sludge from the digesters.

It was hoped that one of the results of removing sludge from the digesters at a high rate would be the dislodging of much of the heavy sludge that is apparently lodged in the sides and corners of the secondary digesters. To aid in getting rid of the heavy sludge, the secondary digester contents were occasionally recirculated by withdrawing from the bottom and pumping back into the middle. At first, some dislodging of heavy sludge occurred at the fast sludge removal rate. After a couple of months, however, the situation began changing drastically. Apparently, the effective volume of the digesters was so reduced by heavy sludge, grit and rags that there was only a very short detention time in the secondary digesters at the high sludge removal rate. The TS concentration of the sludge removed from the digesters decreased from 6.0 percent TS to 3.0 to 3.5 percent TS. Solids-liquid separation was not occurring.

Consequently, the sludge removal rate was cut in half. The plant manager is now planning to hire a private contractor to pump out both secondary digesters, removing the heavy material.

Sludge Treatment and Disposal Costs

In assessing the operational costs of sludge treatment and disposal at Lakewood, the following assumptions were made:

1. All costs are shown as per ton (metric and English) of dry TS removed from the digesters.
2. Costs are based on a digested sludge removal rate of 1,180 t/yr (1,300 ton/yr) in 1974 and 1,810 t/yr (2,000 ton/yr) in 1976.
3. The unit cost factors which were used are:

Item	1977 Costs
Electricity	$0.045/kwh
Fuel oil	$1.454/ℓ ($0.384/gal)
Ferric chloride (40 percent $FeCl_3$)	$0.118/ℓ ($0.031/gal)
Lime	$0.015/kg ($0.033/lb)
Labor categories	
(L) Laborer	$5.00/hr
(R) Repairman	$6.43/hr
(M) Maintenance person	$5.70/hr
(O) Filter-dryer operator	$5.43/hr

The costs of handling the alum sludge were estimated for 1974, when the filter and dryer were operated during one shift per day, and for 1976, when they were operated for two shifts per day. Operation for two shifts was a response to the additional sludge load imposed on the plant by alum addition. Therefore, the additional costs of running for two shifts can be attributed to the alum addition.

The operational cost breakdown is shown in Table C-17. The total treatment costs for each process are summarized in Table C-18. Because of the increase in the number of tons of dry TS removed from the digesters in 1976, the treatment cost per ton of dry TS was not much greater than it was in 1974. The total treatment cost was $133.58/t ($121.19/ton) in 1974 and $138.13/t ($125.28/ton) in 1976.

These costs can be compared with the cost of hauling liquid sludge. The private contractor's fee for hauling the sludge and applying it to agricultural fields with a "field jimmy" is $0.11/ℓ ($0.03/gal). The cost per ton of dry TS of this method depends upon the TS concentration of the liquid sludge. A more concentrated sludge improves the cost-effectiveness of the method. Good digester performance will result in both reduced sludge volume and a more concentrated sludge, which will increase the cost-effectiveness of liquid sludge hauling. When the digesters at Lakewood were producing a 6.0 percent TS sludge, the cost of hauling was $132.30/t ($120/ton). This is less than the treatment using the filters and dryer.

Summary and Conclusions

The Lakewood, Ohio, plant is an older facility (built in 1966) which has treated average annual flows ranging from 42,000 to 55,000 m^3/day (11.2 to 14.6 mgd) during the period examined in this case study. The plant's effluent quality is poor due to hydraulic overloading and inadequate solids handling capacity. The plant treats alum-waste activated and primary sludges which are combined in the gravity thickeners. The basic solids-handling problem facing the plant has been its inability to remove sludge from the digesters at a fast enough rate. The plant is equipped with vacuum filters and a flash dryer for sludge dewatering and drying. Over the last three years, the flash dryer has functioned at loadings of 360 to 450 kg (800 to 1,000 lb) TS/hr. This is 360 to 450 kg/hr (800 to 1,000 lb/hr) less than the dryer's design capability rating. It has been possible to operate only one of the two vacuum filters at a time because of the limiting capacity of the flash dryer.

Alum addition for phosphorus removal placed a squeeze on the plant's already overloaded sludge handling system. It increased the amount of sludge generated, and at the same time decreased the capacity of the vacuum filter and flash dryer to process the sludge. This created a bottleneck in the solids handling end of the plant, leading to high solids buildup and recirculation throughout the plant.

Although no part of the plant was designed specifically for sludge storage, the whole plant has become what could be described as one big sludge storage tank. In addition to the solids buildup in the thickeners and digesters, the thickener overflow and

TABLE C-17. COSTS OF TREATING ALUM SLUDGE DURING SINGLE AND DOUBLE SHIFT VACUUM FILTER AND FLASH DRYER OPERATION, LAKEWOOD, OHIO

	Single Shift Operation (1974)	Double Shift Operation (1976)
Gravity Thickening		
Electricity (kwh/yr)	118,000	118,000
($/t ($/ton))	$4.50 ($4.03)	$2.92 ($2.65)
*Maintenance supplies ($/yr)	$100	$100
($/t ($/ton))	$0.09 ($0.08)	$0.06 ($0.05)
Maintenance labor (hr/yr)	32 (M)	32 (M)
($t ($/ton))	$0.15 ($0.14)	$0.10 ($0.09)
Operating costs and overhead	none	none
Anaerobic Digestion		
Electricity (kwh/yr)	360,000	360,000
($/t ($/ton))	$13.74 ($12.46)	$8.93 ($8.10)
Fuel oil (gal/yr)	1,000	1,000
($/t ($/ton))	$0.32 ($0.29)	$0.21 ($0.19)
*Maintenance supplies ($/yr)	$2,630	$2,630
($/t ($/ton))	$2.23 ($2.02)	$1.44 ($1.31)
Maintenance labor (hr/yr)	832 (M)	832 (M)
($/t ($/ton))	$4.02 ($3.65)	$2.61 ($2.37)
Operational labor (hr/yr)	1,100 (L)	1,100 (L)
($/t ($/ton))	$4.66 ($4.23)	$3.03 ($2.75)
Overhead ($/yr)	$3,230	$3,230
($/t ($/ton))	$2.73 ($2.48)	$1.78 ($1.61)

(continued)

TABLE C-17 (continued)

	Single Shift Operation (1974)	Double Shift Operation (1976)
Chemical Conditioning and Vacuum Filtration		
$FeCl_3$ and lime (\$/t (\$/ton))	\$9.81 (\$8.90)	\$9.81 (\$8.90)
Electricity (kwh/yr) (\$/t (\$/ton))	57,300 \$2.18 (\$1.98)	57,300 \$1.42 (\$1.29)
Maintenance supplies (\$/yr) (\$/t (\$/ton))	\$1,450 \$1.22 (\$1.11)	\$2,200 \$1.21 (\$1.10)
Maintenance and repair labor (hr/yr) (\$/t (\$/ton))	792 (M,R) \$4.02 (\$3.65)	1,200 (M,R) \$3.97 (\$3.60)
Operational labor (hr/yr) (\$/t (\$/ton))	830 (0) \$3.81 (\$3.46)	2,490 (0) \$6.89 (\$6.25)
Overhead (\$/yr) (\$/t (\$/ton))	\$2,920 \$2.48 (\$2.25)	\$6,225 \$3.43 (\$3.11)
Flash Drying		
Electricity (kwh/yr) (\$/t (\$/ton))	326,500 \$12.46 (\$11.30)	609,500 \$15.12 (\$13.71)
*Fuel oil (\$/yr) (\$/t (\$/ton))	\$42,900 \$36.38 (\$33.00)	\$77,000 \$42.45 (\$38.50)
*Maintenance supplies (\$/yr) (\$/t (\$/ton))	\$3,036 \$2.57 (\$2.33)	\$4,600 \$2.54 (\$2.30)
Maintenance labor (hr/yr) (\$/t (\$/ton))	1,188 (M,R) \$6.04 (\$5.48)	1,800 (M,R) \$5.95 (\$5.40)
†Operational labor (hr/yr) (\$/t (\$/ton))	914 (L) 1,240 (0) \$9.59 (\$8.70)	1,830 (L) 3,730 (0) \$15.37 (\$13.94)
Overhead (\$/yr) (\$/t (\$/ton))	\$5,830 \$4.94 (\$4.48)	\$12,220 \$6.74 (\$6.11)

(continued)

TABLE C-17 (continued)

	Single Shift Operation (1974)	Double Shift Operation (1976)
Flash Drying (cont'd)		
Approximate average income--dried product ($/t ($/ton))	$5.64 ($5.12)	$2.15 ($1.95)

[*] 1976 costs.

[†] Includes the cost of bagging and trucking the dried sludge.

TABLE C-18. COSTS OF ALUM SLUDGE TREATMENT DURING SINGLE AND DOUBLE SHIFT VACUUM FILTER AND FLASH DRYER OPERATION, LAKEWOOD, OHIO

	Single Shift Operation (1974)	Double Shift Operation (1976)
Gravity thickening:	$4.74 ($4.30)	$3.08 ($2.79)
Anaerobic digestion:	$27.70 ($25.13)	$18.00 ($16.33)
Chemical conditioning and vacuum filtration:	$23.52 ($21.35)	$26.73 ($24.25)
Flash drying:	$77.62 ($70.41)	$90.32 ($81.91)
Total:	$133.58 ($121.19)	$138.13 ($125.28)

digester supernatant sidestreams carried heavy solids loads back to the head of the plant. The solids accumulated in the primary clarifiers from which they were fed back to the thickeners and digesters, thus being trapped essentially in a closed-loop system. Furthermore, because the thickeners were overloaded, 97 to 98 percent of the secondary sludge which was produced was recirculated through the aeration basins and back to the final clarifiers. The solids accumulated in the secondary clarifiers, and then were fed back to the aeration basins, forming a second closed loop.

An additional 2.3 t/day (2.5 ton/day), or 4.3 t/mil m^3 (360 lb/MG), of secondary sludge were wasted to the thickeners during alum addition (Figures C-11 and C-12). The volume of secondary sludge wasted did not increase, however. The volume of thickened sludge pumped to the digesters also did not change significantly, but the sludge TS concentration increased from 4.3 to 5.6 percent. The increase in the mass of sludge pumped to the digesters was approximately the same as the amount of additional secondary sludge wasted to the thickeners.

During alum addition, the sidestreams from the sludge treatment operations contributed a loading of 6,948 kg SS/day (15,300 lb SS/day) to the primary clarifiers. This figure is greater than the 6,025 kg/day (13,270 lb/day) which was contained in the plant influent. By far the most significant sidestream in terms of solids input was the thickener overflow, carrying 6,433 kg SS/day (14,170 lb SS/day). Alum addition increased the overloading of the thickeners, and problems with bulking and poor overflow quality became more severe.

The rate at which sludge could be flash dried was reduced during alum addition because the moisture content of the filter cake rose from 76.2 percent to 78.6 percent. Filter yield was then cut back so that it would not exceed the drying rate. The volume of sludge removed from the digesters each day was decreased in response to the reduced filter and dryer capacity. This resulted in the production of 22.7 m^3/day (6,000 gal/day) more digester suprenatant. The digesters were overloaded to the extent that the supernatant TS concentration rose from 2.4 to 3.1 percent. The supernatant TS loading on the primary clarifiers increased by 1,362 kg TS/day (3,000 lb/day).

Alum addition to the activated sludge effluent channel affected wastewater treatment operations in several ways:

- The primary clarifier influent concentrations of SS and BOD were increased due to greater loadings from the thickener overflow and digester supernatant.
- The primary clarifier effluent concentrations of SS and BOD did not increase, however, indicating that the additional loadings were removed as sludge.

- The MLSS concentration in the aeration basins rose by 56 percent due to a higher TS concentration of the return activated sludge.

- The capture of SS and BOD during secondary clarification was increased, preventing an increase in final effluent SS and BOD_5 concentrations.

- Phosphorus removal was increased from 20 percent to only 65 percent. Greater removal could have been achieved at higher alum dosages or with polymer addition, but the plant was not equipped to handle the additional sludge which would have been generated.

In order to remove sludge from the digesters at a faster rate and thus alleviate the solids-handling problem, the length of filter and dryer operation was extended to 14 hr/day instead of 6.5. A reduction in the volume of supernatant and improved supernatant quality resulted. More recently, the plant returned filter and dryer operation to one shift per day and began hauling liquid sludge. Initially, during liquid sludge hauling, the rate of sludge removal from the digesters exceeded the feed rate by 908 kg TS/day (2,000 lb TS/day). As a result, the volume of supernatant was further reduced and there was more room in the secondary digesters for solids-liquid separation.

The indirect effects of longer filter-dryer operation and liquid sludge hauling included less primary sludge production, less frequent thickener bulking, a reduction of the MLSS concentration in the aeration basins, and improved plant final effluent quality.

CASE STUDY G: MENTOR, OHIO

Introduction

The Greater Mentor Wastewater Treatment Plant provides an example of secondary addition of alum for phosphorus removal. The chemical sludge at the plant is handled separately from the primary sludge. The alum-secondary sludge wasted from the secondary clarifiers is sent to an aerobic digester followed by dual cell gravity concentrators (DCG's) with final disposal on crop lands. Wastewater influent to the plant is delivered by separate sanitary sewers and contains approximately 12 percent industrial waste.

History

The history of modifications to the plant which have affected sludge, follows:

- Late 1965 - Start-up of primary plant

- Early 1969 - Start-up of ferrous iron and lime addition for phosphorus removal
- October 6, 1973 - Start-up of secondary treatment using contact stabilization activated sludge process
- October 7, 1973 - End of ferrous iron and lime addition
- October 14, 1974 - Start-up of alum addition for phosphorus removal
- June 22, 1976 - Switch from contact stabilization to conventional activated sludge process
- March 25, 1977 - Return to contact stabilization activated sludge process
- July 1977 - Switch from aluminum sulfate to sodium aluminate addition for phosphorus removal.

Two periods were selected for comparison purposes to determine the impact of chemical additions for phosphorus removal. The period from mid-October 1973 to mid-October 1974 was selected as a period without phosphorus removal. The period from July 1976 to March 1977 was selected as a period during which phosphorus removal with alum was practiced. Table C-19 presents the daily average influent and effluent wastewater characteristics and plant removal efficiencies during the two periods.

TABLE C-19. INFLUENT AND EFFLUENT WASTEWATER CHARACTERISTICS AND REMOVAL EFFICIENCIES, MENTOR, OHIO

		October '73 to October '74	July '76 to March '77
Plant Flow mil m^3/day (mgd)		0.017 (4.60)	0.02 (5.29)
SS (mg/ℓ)	Primary Influent	186	172
	Primary Effluent	113	91
	Plant Effluent	36	35
	% Removal - Primary	39	47
	% Removal - Plant	81	80
BOD (mg/ℓ)	Primary Influent	205	187
	Primary Effluent	147	137
	Plant Effluent	32	26
	% Removal - Primary	29	27
	% Removal - Plant	84	86
TP (mg/ℓ)	Primary Influent	.28	25
	Plant Effluent	21	14
	% Removal - Plant	25	44

Chemical Addition for Phosphosrus Removal

Liquid aluminum sulfate (9.0 percent Al^{+3}) was added to the effluent channel leading from the aeration basins to the final clarifiers while the plant was operated as a conventional activated sludge system with four aeration basins. Approximately 2.6 m^3 (690 gal) of liquid alum were used per day. The actual feed rate of solution was changed twice a day to account for diurnal variations in plant flow. From 7 a.m. to 5 p.m. the solution was added at a rate which achieved a dosage of 12.2 mg/ℓ Al^{+3}, with a plant flow of about 0.017 mil m^3/day (4.5 mgd). At 5 p.m. the chemical feed rate was changed to maintain the same chemical dosage at a plant flow of about 0.024 mil m^3/day (6.5 mgd).

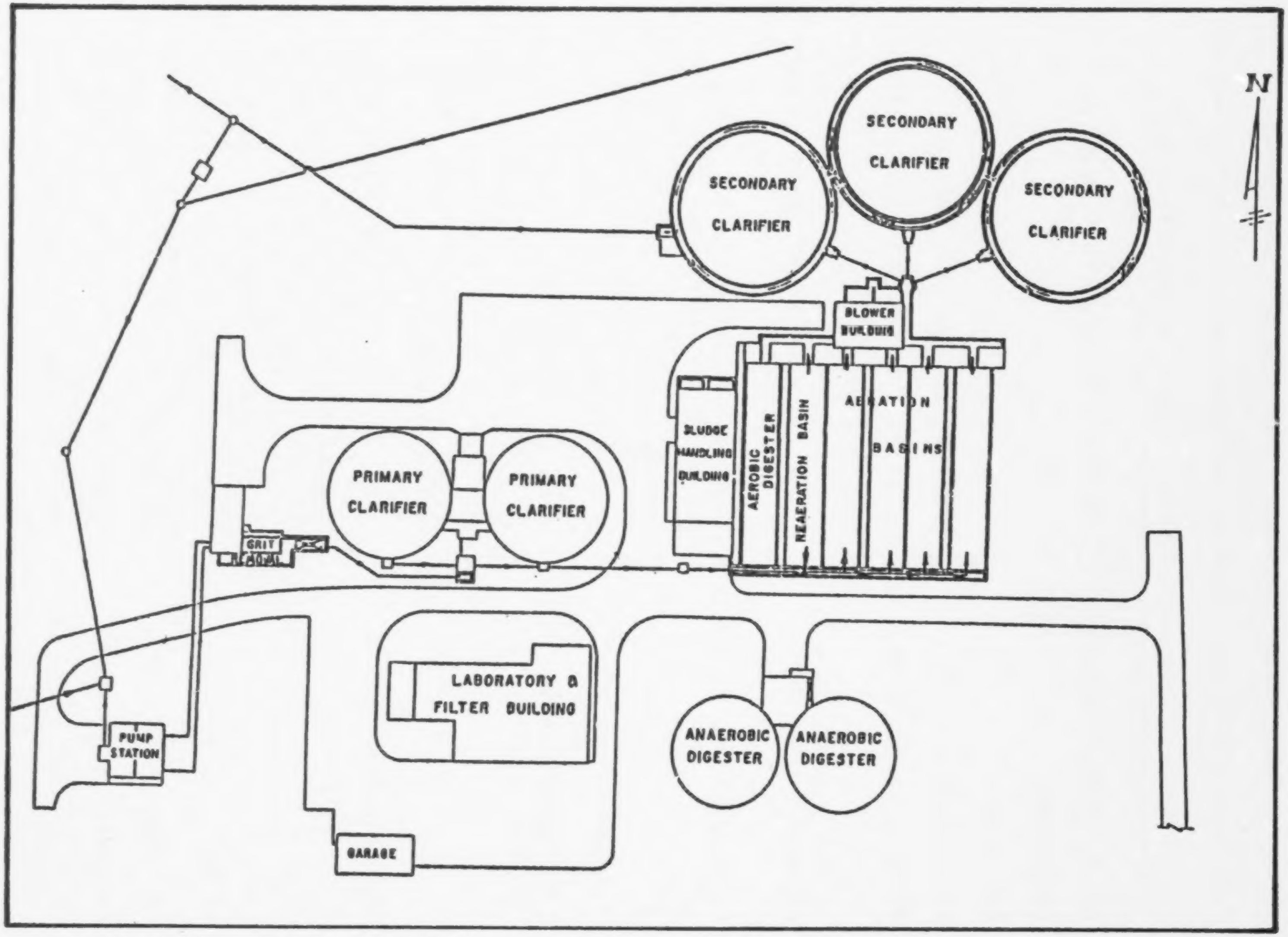
In addition to alum, the plant has used several other chemicals for phosphorus removal. Initially, ferrous iron and lime were added ahead of the primary clarifiers during the period when the plant only provided primary treatment. More recently, sodium aluminate addition ahead of the aeration basins was tested. According to plant personnel, sodium aluminate addition caused the pH of the sludge in the aerobic digester to increase significantly. This, in turn, decreased the effectiveness of the DCG's. As a result, the use of sodium aluminate was abandoned.

General Description of Wastewater Treatment Operations Affecting Sludge

A wastewater flow diagram is presented in Figure C-15. Table C-20 is a summary of the wastewater and sludge handling unit and mechanical equipment. Influent wastewater flows to the pump station, where it passes through either a comminutor or a bar screen before dropping into the raw wastewater wet well. The wastewater is then pumped to the aerated grit chamber where heavy, abrasive solids are removed. The wastewater then flows to the primary clarifiers where settleable and floating solids are removed.

The primary effluent flows by gravity to the aeration basins where it is mixed with return sludge. From the aeration basins the mixed liquor flows to the secondary clarifiers where the sludge and solids in the wastewater are settled out. The secondary effluent is chlorinated and discharged to Lake Erie. A portion of the sludge from the secondary clarifiers is returned to the reaeration basins and is subsequently recycled by mixing with the aeration basin influent.

Sludges removed for sludge processing from the primary and secondary treatment systems are handled separately. Primary sludge is pumped to the anaerobic digesters or to the heat treatment unit. Digested sludge is then chemically conditioned and sent to the vacuum filters. Thermally conditioned sludge from the heat treatment unit is sent without chemical conditioning to another set of vacuum filters. Vacuum filter sludge cake is disposed of on croplands.



N
SECONDARY CLARIFIER
SECONDARY CLARIFIER
SECONDARY CLARIFIER
BLOWER BUILDING
AERATION BASINS
REAERATION BASIN
AEROBIC DIGESTER
SLUDGE HANDLING BUILDING
PRIMARY CLARIFIER
PRIMARY CLARIFIER
GRIT REMOVAL
LABORATORY & FILTER BUILDING
ANAEROBIC DIGESTER
ANAEROBIC DIGESTER
PUMP STATION
GARAGE

Figure C-15. Greater Mentor wastewater treatment plant wastewater flow diagram.

TABLE C-20. GENERAL PLANT DESCRIPTION SUMMARY, MENTOR, OHIO

Unit	No. and Description
Plant Design Capacity	0.045 mil m^2/day (12.0 mgd).
Comminutor	
Bar Screen	
Grit Chamber	78.4 m^3 (2,800 ft^3) aerated.
Primary Clarifiers	2 at 21.3 m (70 ft) dia. x 2.7 m (9 ft); overflow weir - 2 at 193.5 m (635 ft) length.
Primary Sludge Pumps	2 at 7.5 hp; 0.47 m^3/min (125 gal/min) capacity.
Primary Sludge Collectors	2 at 1.6 revolutions/hr, sludge scrapers on the bottom of a rotating arm direct the sludge on the bottom of each clarifier to the hopper in the center of the tank; motors - 2 at 1/2 hp.
Influent Flume to Aeration Basins	48.5 m (159 ft) length x 1.1 m (3.5 ft) width x 0.9 m (3 ft) effective dia.; 3 2-in air diffuser headers at 17-20 diffusers each capable of discharging up to 12 cfm at 10.6 in water.
Aeration Basins	6 at 37.5 m (123 ft) length x 7.3 m (24 ft) width x 4.6 m (15 ft) dia. 3 multistage centrifugal compressors (blowers) at 300 hp each capable of supplying 5600 scfm at 7 psig; 5 4-in diffuser headers at 32 diffusers per basin, each capable of discharging up to cfm 26.9 cm (10.6 in) water; aeration basin #6 is used as an aerobic digester.
Secondary Clarifiers	3 at 28 m (92 ft) dia. x 2.4 m (8 ft) SWD; overflow weir 3 at 88.1 in (289 ft) length.

(continued)

TABLE C-20 (continued)

Unit	No. and Description
Secondary (Return) Sludge Pumps	2 at 25 hp, 7.6 m^3/min (2,000 gal/min); 1 at 40 hp, 15.9 m^3/min (4200 gal/min).
Secondary Sludge Collector	3 at 1.8 revolutions/hr, sludge scrapers on bottom of pair of rotating arms with 5 6-in "suction" pipes per arm; motors - 3 at 3/4 hp; scrum removal by travelling weir attached to torque tube of scraper arms.
Anaerobic Digesters	2 at 1719.8 m^3 (61,420 ft^3), recirculation pumps - 2 at 0.57 m^3/min (150 gal/min); heat exchanger boiler - 1 at 500,000 BTU/hr.
Vacuum Filter (for anaerobic digester sludge)	1.8 to 3.6 kg (4 to 8 lb/hr) loading rate at 30 percent dry solids.
Sludge Heat Treatment (for raw primary sludge)	3.8 m^3/hr (1,000 gal/hr) feed rate at 5 percent dry solids. Sludge temperatures °C (°F). Inlet 12.7 (55) Maximum 202.4 (400) Outlet 37.4 (100) Disintegrators 2 at 30,280-94,625 mil m^3/min (8-25 gal/min) of 3-10% solids. High pressure pumps 2 at 49,205 75,700 mil m^3/min (13-20 gal/min) at 300 psig. Head exchanger, 3.03-4.54 m^3/hr (800-1200 gal/hr). Reactor, 3.03-4.54 m^3/hr (800-1,200 gal/hr) at 275 psig and 210°C (414°F).

(continued)

TABLE C-20 (continued)

Unit	No. and Description
	Decant and Thickening Tanks
	Solids loading - 97.7 kg/m^2/day (20 lb/ft^2/day) maximum.
	Sludge pumps - 2 at 18,925-180,250 mil m^3/min (5-50 gal/min).
	Vacuum filter (for heat) treatment sludge) - 302.8 kg/hr (667 lb/hr) loading rate at 30 percent dry solids.
Sludge Holding Tank (Thickener)	84.4 m^3 (22,300 gal) capacity.
Dual Cell Gravity (DCG) Concentrator	3 at 11.35 m^3/hr (3,000 gal/hr) total capacity (aerobically digested sludge); main drives - 3 at 1/2 hp; internatl conveyors - 3 at 1½ hp, 110 v.; external conveyors - 2 at 2 hp, 220 v., 61 cm (24 in) width, capacity - 2.8 m^3 (100 ft^3) sludge at 1,297 kg/m^3 (80 lb/ft^3); sludge pumps - 3 adjustable speed and stroke plunger type at 2 hp, 440 v., 26,495 to 211,960 mil m^3/min (7 to 56 gal/min); sludge pump wet wells - 1 at 33.1 m^3 (8,740 gal), 1 at 41.6 m^4 (10,980 gal); chemical mixing - 4 plastisol lined steel tanks at .38 m^3 (100 gal), with 1/4 hp mixers operating at 1,725 rpm; chemical feed pumps - 3 dual head at 1/4 hp, 75,700 mil m^3/hr (20 gal/hr) head; coagulation - 3 tanks; filter media - fine mesh, nylon cloth.
Lime Clarification of Supernatants and Filtrates	6.7 m (22 ft) length x 6.7 m (22 ft) width x 3.7 m (12 ft) dia.; lime storage bin - 3.7 m (12 ft) dia. x 6.5 SWD; maximum lime feed rate .25 m^3/hr (8.75 ft^3/hr; lime sludge pumps - 2 at 0.19 m^3/min (50 gal/min).

Alum-secondary sludge is pumped to a sludge holding tank and on to the aerobic digester, or is sent directly to the aerobic digester. Following aerobic digestion, the sludge is dewatered by the DCG's. Dewatered sludge is disposed of at a privately owned dumpsite. Aerobic digester and sludge holding tank decantate and DCG concentrate are returned to the head of the aeration basins.

The plant also has a lime clarifier in which anaerobic digester supernatant, vacuum filter filtrate, DCG concentrate, and sludge holding tank decantate can be treated to reduce phosphorus concentrations. At present, the lime clarifier is not operated on a continuous basis. The anaerobic digester supernatant and vacuum filter filtrate are returned to the head of the primary clarifiers.

Detailed Description of Wastewater Treatment Operations Affecting Sludge

Since the alum-secondary sludge is handled separately from primary sludge, this section will emphasize plant operations affecting secondary sludge generation and handling.

Primary Clarifiers--

Since alum addition was initiated, the quantity and characteristics of the primary clarifier effluent have changed as previously indicated in Table C-19. Specifically, there has been a noticeable improvement in primary clarifier performance in terms of suspended solids removals. This has led to a decrease in solids loadings on the secondary treatment system. At the same time, concentrations of BOD's discharged from the primary clarifiers have also decreased. Plant data does not, however, fully account for slugs of concentrated waste from septic tank pumpouts delivered to the treatment plant by truck. During the period without phosphorus removal, two to six 3.78-m^3 (1,000-gal) loads entered the system daily. This practice was terminated prior to alum addition.

Aeration Basins--

During the period without phosphorus removal, the aeration basins were operated as a contact stabilization system with two contact basins and one reaeration basin. At that time, the contact basins contained 2,400 to 2,800 mg/ℓ MLSS, with approximately 70 percent volatile solids, and the reaeration basin was operated at 5,000 to 7,000 mg/ℓ MLSS with roughly the same percent volatile solids as the contact basins.

During the studied period with aluminum sulfate addition the system was operated as a conventional activated sludge treatment plant with four aeration basins, operated at concentrations of

3,000 to 7,500 mg/ℓ MLSS, with 62 percent volatile solids. Return sludge with 6,000 to 20,000 mg/ℓ suspended solids was recycled at a rate equal to 40 to 50 percent of plant forward flow. This led to considerably longer sludge ages of the mixed liquor in the aeration basins, causing the cells to enter an endogenous growth phase. Less sludge was wasted per pound of BOD removed in the secondary treatment system during this period than during the period without phosphorus removal.

Secondary Clarifiers--

Sludge is removed from the bottoms of the secondary clarifiers on a continuous basis by the sludge draw-off pipes attached to the arms of the sludge collector mechanisms. This sludge is returned to the aeration basin system at a pumping rate that varies between 25 and 50 percent of the plant forward flow. For approximately 1 hr/8-hr shift, sludge is wasted to sludge processing at a pumping rate approximately equal to 100 percent of a plant forward flow. This high rate of pumping is necessary to remove the sludge blanket which accumulates to up to 1.8 m (6 ft) at the center of the clarifier during periods of low rate return sludge pumping.

Aerobic Digester--

During the period prior to alum addition, one of the six aeration basins was used as an aerobic digester. Generally, during the period of alum addition, one aeration basin was used as an aerobic digester, although two aeration basins were used as aerobic digesters for a short time. Characteristics of the sludge influent to and effluent from the aerobic digester are given in Table C-21.

TABLE C-21. AEROBIC DIGESTER SLUDGE CHARACTERISTICS, MENTOR, OHIO

	October '73 to October '74	July '76 to March '77
Influent Sludge:		
WAS kg (lb) dry solids per month	38,600 (85,000)	45,400 (100,000)
Total Solids (%)	0.5-1.5	0.8-2.0
Volatile (%)	70	63
Effluent Sludge (to DCG Concentrators):		
Total Solids (%)	2.0-2.6	2.4-2.9
Volatile (%)	68	56

The aerobic digesters are aerated and mixed by aeration equipment identical to that in all other aeration basins at the Mentor plant. Sludge in the digesters is maintained at approximately 27°C by the diffused air which is heated by the air compressors. Supernatant is decanted from the top of the aerobic digester less than once per day. Prior to drawing off supernatant, air to the tank is turned off and sludge wasting discontinued for 2 to 6 hr. After settling has taken place, supernatant is withdrawn slowly to prevent mixing in the tank. During supernatant withdrawal, concentrations of dissolved oxygen in the tank drop as low as 0.1 mg/ℓ. When decanting is complete, air is turned on again and aeration takes place until the dissolved oxygen levels reach 4 to 5 mg/ℓ. This takes approximately 1 day, at which point the tank is again ready for decanting.

Normally, after 1 to 2 wk of continuously wasting sludge into the tank and drawing off supernatant, the sludge in the tank becomes so thick that no additional water can be decanted off. At this time sludge must be pumped to the DCG's. To avoid this situation, the DCG's are operated daily if at all possible.

The impacts of chemical additions for phosphorus removal on aerobic digester operations have been an increase in the concentration of total solids fed to the digester coupled with a decrease in the fraction of those solids which are volatile. The destruction of volatile solids by the aerobic digester was also observed to increase after alum addition to the secondary treatment system was initiated.

Dual Cell Gravity Concentrators--

Three DCG's are located in the sludge handling building. These units are used to dewater the aerobically digested sludge prior to land disposal of wet sludge cake. Chemical storage tanks, pumps, and piping are provided for adding polymers to the sludge ahead of the mixing tank as previously detailed in Table C-20.

Figure C-16 presents a schematic diagram of the DCG's and appurtenances. As shown in the diagram, sludge from the aerobic digesters enters one of two sludge wet wells prior to pumping to the DCG units. Conditioning chemicals may be injected into the sludge either immediately following the sludge pumps or immediately preceding the coagulation tanks.

The chemical feed system consists of three chemical feed pumps and four chemical storage tanks. Four tanks are furnished in case both anionic and cationic polymers are required for coagulation. Sludge enters the bottom of coagulation tanks, mixes with coagulation chemicals, and overflows through four outlets near the top of each tank.

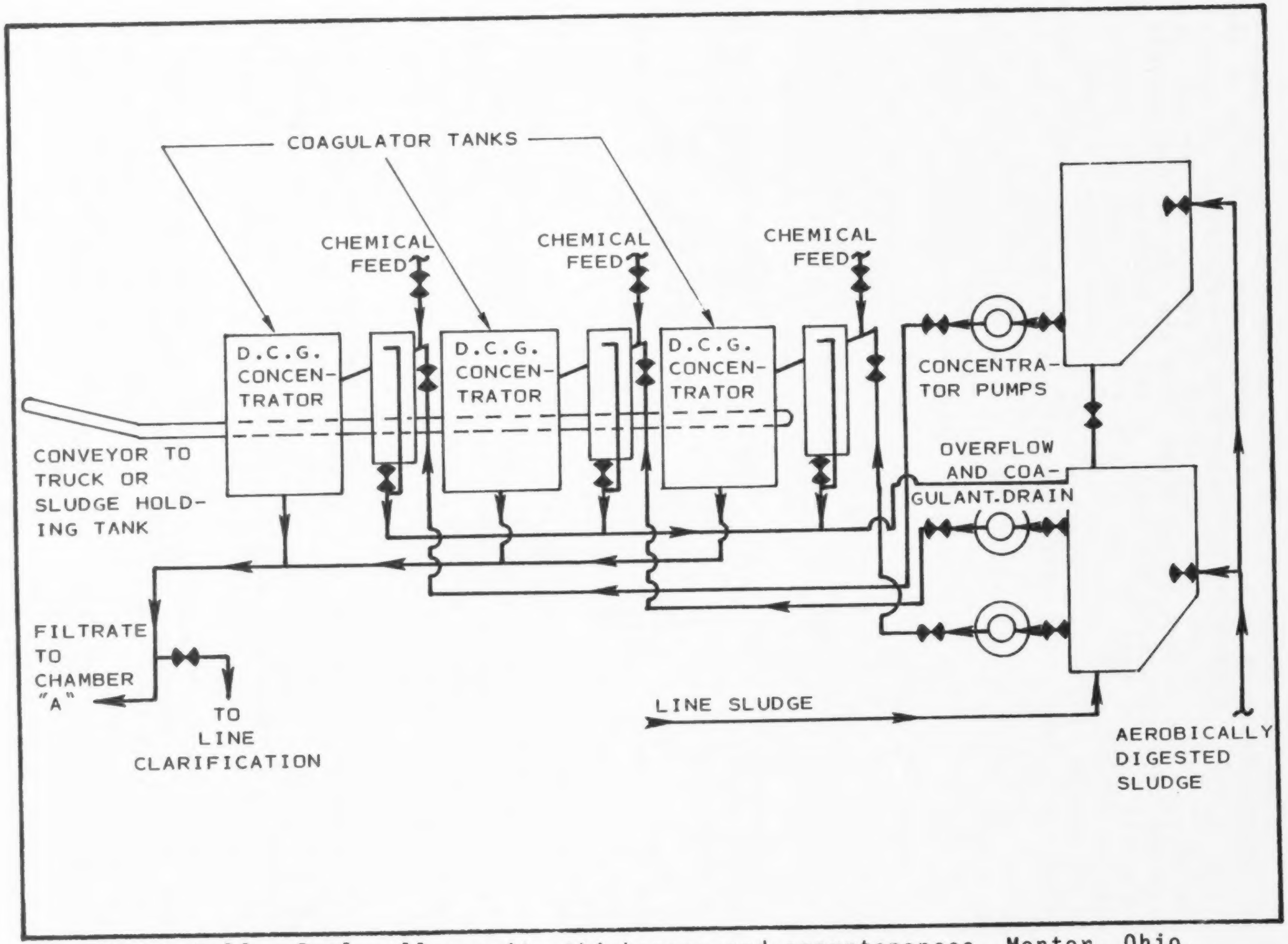

COAGULATOR TANKS
CHEMICAL FEED
CHEMICAL FEED
CHEMICAL FEED
D.C.G. CONCENTRATOR
D.C.G. CONCENTRATOR
D.C.G. CONCENTRATOR
CONCENTRATOR PUMPS
OVERFLOW AND COAGULANT.DRAIN
CONVEYOR TO TRUCK OR SLUDGE HOLDING TANK
FILTRATE TO CHAMBER "A"
TO LINE CLARIFICATION
LINE SLUDGE
AEROBICALLY DIGESTED SLUDGE

Figure C-16. Dual cell gravity thickeners and appurtenances, Mentor, Ohio.

Sludge slurry from the coagulation tanks enters the first cell of each of the dual cell concentrators at four points. The cells are formed by a moving, fine mesh nylon filter cloth. A hump in the center of the cloth divides each unit into a thickening cell and a compression cell. Coagulated solids entering the first or thickening cell are trapped on the cloth and carried over the hump to the compression cell while the filtrate drains through the cloth. In the compression cell, the solids form a plug, or cake roll, which in turn presses additional moisture out of the sludge. Excess quantities of sludge cake are discharged over the rim of the second cell onto one of the internal conveyor belt system. This system is comprised of one horizontal conveyor belt which receives the sludge from the internal conveyors and discharges the sludge cake to the second, inclined conveyor belt. The second conveyor belt extends outside the sludge handling building and discharges the sludge cake to a waiting dump truck.

Normally, three DCG units are utilized 90 percent of the time and two DCG units are used 10 percent of the time. The units are not operated at all unless at least two units are in working order. During the winter months, subfreezing weather often prevents conveyor system operation, even with the actual DCG units in good repair. Lack of trucks available for hauling wet sludge cake can also prevent DCG operation. Specific mechanical problems which occur in the DCG system were identified by plant personnel as follows: screens and mesh tear and develop holes; zippers break from seals or loose teeth; seals wear, break, crack, or get caught in drive sprockets; conveyor bearings require replacement; polymer pumps clog due to the formation of scale; main drive unit and conveyor motors burn out.

According to plant personnel, the DCG units are operated as often as possible. Although no operation or maintenance problems have been directly attributable to the handling of alum sludge, the units have been operated somewhat less frequently since the initiation of alum addition. The difference appears negligible, however, if the fact is considered that the units were operated less than 45 percent of the available hours per month during both periods.

Performance of the DCG units is shown in Table C-22. Since the initiation of alum addition, the following operational changes have taken place:

- The volume of sludge fed per hour of operation has decreased slightly.
- The mass of dry solids fed to the DCG's per hour of operation has increased slightly.
- The amount of cationic polymer added per gallon of sludge fed to the DCG's has increased slightly (although the

TABLE C-22. PERFORMANCE OF DUAL CELL GRAVITY CONCENTRATORS, MENTOR, OHIO

Estimated Monthly Averages	Without P-Removal October '73 to October '74	With P-Removal July '76 to March '77
Hours of operation	330	315
Percent TS feed sludge (including conditioners)	2.1-2.7	2.5-3.1
m^3 (gal)feed sludge/hr of operation	3.7 (980)	3.5 (920)
kg (lb) feed sludge TS/hr of operation	77.2-99.9 (170-220)	86.3-109 (190-240)
kg (lb) cationic polymer/m^3 (1000 gal) feed sludge	1.7 (14.4)	1.9 (15.9)
kg (lb) anionic polymer/mil m^3 (1000 gal) feed sludge	473 (0.0394)	708 (0.0059)
kg (lb) wet sludge cake/mo	299,000 (658,000)	345,000 (760,000)
Percent TS sludge cake	8.8-9.2	8.2-9.1
kg (lb) cake TS/mo	26,800 (59,000)	30,900 (68,000)
kg (lb) cake TS/MG plant influent	1.54 (12,830)	1.54 (12,850)

amount added per pound of solids fed to the DCG's has remained the same).

- The amount anionic polymer added per gallon of sludge fed to the DCG's has decreased considerably.
- The concentration of total solids in the wet sludge cake removed from the DCG's has decreased marginally.
- Overall, the mass of TS concentrated per million gallons of plant influent has remained the same.

Summary and Conclusions

Since the initiation of alum addition for phosphorus removal, no increase in the mass of secondary sludge TS removed from the plant by DCG units has been observed. This unexpected result can be explained by a combination of several factors. These factors, and other impacts of chemical addition for phosphorus removal on the plant, are highlighted below.

First, there has been a general decrease in the loadings of SS and BOD_5 influent to the secondary treatment system. This has been the result of both improved primary clarifier performance and the termination of septic tank pump-out deliveries to the plant. In fact, the actual magnitude of the decrease due to termination of septic tank waste deliveries is somewhat greater than indicated by available plant data. As a result of plant composite sampling which takes place once every 3 hr, 24 hr/day, slugs of septic tank wastes within the primary treatment system often go largely undetected.

Second, the aeration system was fun as a contact stabilization process during the period without phosphorus removal. During the period of alum addition for phosphorus removal, the plant was run as a conventional activated sludge plant. Operationally, the aeration system showed significant increases in MLSS concentrations, decreased percentages of volatile solids, and longer sludge ages. This had led to a reduction of the mass of sludge wasted to the aerobic digester per pound of BOD removed.

There have been several other changes observed in plant operations since alum addition was begun, although they are not necessarily atrributable to the alum addition. Specifically, the average concentration of solids in sludge entering the aerobic digester has increased, the volatile fraction of these solids has been reduced, and there appears to be greater volatile destruction taking place in the digester.

Finally, initiation of alum addition, changes in wastewater characteristics, and variations in plant operations have combined to produce several minor changes in DCG concentrator operators.

The volume of sludge fed to the DCG units per hour of operation has decreased slightly, but the mass of TS fed has increased somewhat. The necessary dosage of cationic polymer used to condition the sludge fed to the DCG's increased slightly, while the dosage of anionic polymer used decreased. Overall, the TS concentration in the sludge cake decreased marginally, and the mass of dry cake produced per million gallons of plant influent remained approximately the same. Thus, alum addition has had only minor impacts on plant operations.

CASE STUDY H: BROOKFIELD, WISCONSIN (FOX RIVER)

Introduction

The Brookfield plant provides an example of secondary addition of ferrous sulfate for phosphorus removal. It also provides an example pf pressure (plate and frame) filtration and multiple hearth incineration of an iron chemical sludge which is composed of combined primary and waste activated sludges. Incinerator performance and the impact of ferrous sulfate addition can be compared with other plants (such as Sheboygan, Wisconsin) which operate fluidized bed incinerators.

Raw wastewater influent to the Brookfield plant contains no industrial wastes. Monthly averages of daily influent flows range from 6,430 to 15,100 m^3/day (1.7 to 4.0 mgd) with an average of approximately 9,080 m^3/day (2.4 mgd). The plant was designed to handle an average daily flow of 18,900 m^3/day (5.0 mgd) and thus currently is operating at approximately half capacity.

Table C-23 presents plant influent and effluent wastewater characteristics and removal efficiencies for the periods selected for comparison before and after the initiation of chemical addition for phosphorus removal. Table C-23 indicates that the effluent suspended solids concentrations for the two periods were quite different. However, the influent wastewater sampling point actually includes sidestream flows returned to the head of the plant. Thus, according to plant personnel, the raw influent wastewater characteristics remained unchanged with the difference due to an increase in sidestream suspended solids.

History

Construction of the Brookfield plant was begun in August 1971, with wastewater treatment beginning on January 2, 1974. The following shows the history of modifications to the plant affecting sludge production and characteristics.

January 1974 - Wastewater treatment operations begun using the contact stabilization activated sludge process.

February 1976 to July 1976 - Two final clarifiers in operation (a single final clarifier was in operation during all other periods).

TABLE C-23. INFLUENT AND EFFLUENT WASTEWATER CHARACTERISTICS AND REMOVAL EFFICIENCIENCIES, BROOKFIELD, WISCONSIN

	After P-removal initiated Aug. and Sept. '76, Jan., Feb., & May '77	Before P-removal initiated Jan., Feb., May, Aug., & Sept. '75
Flow m^3/d (mgd)	8,250 m^3/d (2.18 mgd)	9,580 m^3/d (2.53 mgd)
SS (mg/ℓ) -		
influent	265*	212
effluent	18	13
% removal	93	94
BOD (mg/ℓ) -		
influent	113*	103
effluent	7	20[†]
% removal	94	81

*Includes sidestreams returned to lift station at head of plant.

[†] In 1976, effluent BOD concentrations improved when polishing lagoon volume was decreased, reducing retention times from approximately 12 days down to 7 hours. Thus, the apparent decrease in effluent BOD concentration after phosphorus removal was initiated was due to the size reduction of the polishing lagoon.

March and April 1975 - Chemical additions for phosphorus removal were tested, using several different chemicals.

June 1976 - Chemical addition for phosphorus removal using ferrous sulfate (pickle liquor) was initiated.

Chemical Addition for Phosphorus Removal

Pickle liquor, or liquid ferrous sulfate (12.4 percent Fe), is added to the activated sludge mixed liquor at a point 3/4 of the distance between the head end and the discharge end of the contact basin. The quantity of pickle liquor to be added is determined by testing the phosphorus concentration in the effluent from the primary clarifiers once a week. During periods of low flow, the chemical delivery pumps are adjusted to deliver approximately 0.908 kg (2.0 lb) of iron by weight per kg (lb) of phosphorus in the primary effluent. During periods of high flow, the pumps are adjusted to deliver approximately 0.77 kg (1.7 lb) of iron per kg (lb) of phosphorus. Approximately 246 m^3 (65,000 gal) of pickle liquor are added each year, at a rate of 0.673 m^3 (178 gal) of liquid or 0.70 m^3 (184 lb) Fe/day.

General Description of Wastewater and Sludge Treatment Operations

Figure C-17 presents a flow diagram of the Brookfield treatment plant. Table C-24 is a summary of the specifications of the major components of the treatment system. The following paragraphs describe the general sequence of wastewater treatment operations at the Brookfield plant.

Raw influent wastewater passes through a comminutor and is discharged to the plant influent lift station wet well. At the lift station wet well the raw wastewater mixes with sidestreams from several plant operations, including pressure filter filtrate, incinerator scrubber water, digester supernatant, and waste activated sludge. From the lift station, the raw wastewater and returned sidestreams are pumped to the primary clarifiers where primary sedimentation takes place. The primary effluent flows by gravity to the preaeration chamber followed by the contact basin, where chemical addition for phosphorus removal takes place. The mixed liquor then passes to the final clarifier where the overflow is chlorinated and discharged.

Sludge from the bottom of the final clarifier is pumped to the reaeration basin as return activated sludge or is wasted to one of several other locations. In general, waste activated sludge is pumped to the aerobic digester or returned to the lift station wet well at the head of the plant. Sludge from the aerobic digester is subsequently transferred to the sludge processing holding tank. Sludge from the primary clarifiers is generally pumped to the sludge processing holding tank or the aerobic digester.

Sludge received by the sludge processing holding tank is fed along with chemical conditioners to the plate and frame pressure filter for dewatering. Dewatered sludge cake is either incinerated in the multiple hearth furnace or hauled away from the plant as wet sludge cake for disposal on crop land or at a landfill. Incinerator ash is used as a sludge conditioner prior to pressure filtration, or hauled to a landfill for disposal.

Detailed Description of Wastewater and Sludge Treatment Operations

Figures C-18 and C-19 present materials balances for primary and secondary wastewater treatment operations at Brookfield. Data presented with a subscript letter A are for the period after phosphorus removal was initiated, while data with the subscript letter B are for the period before phosphorus removal was initiated.

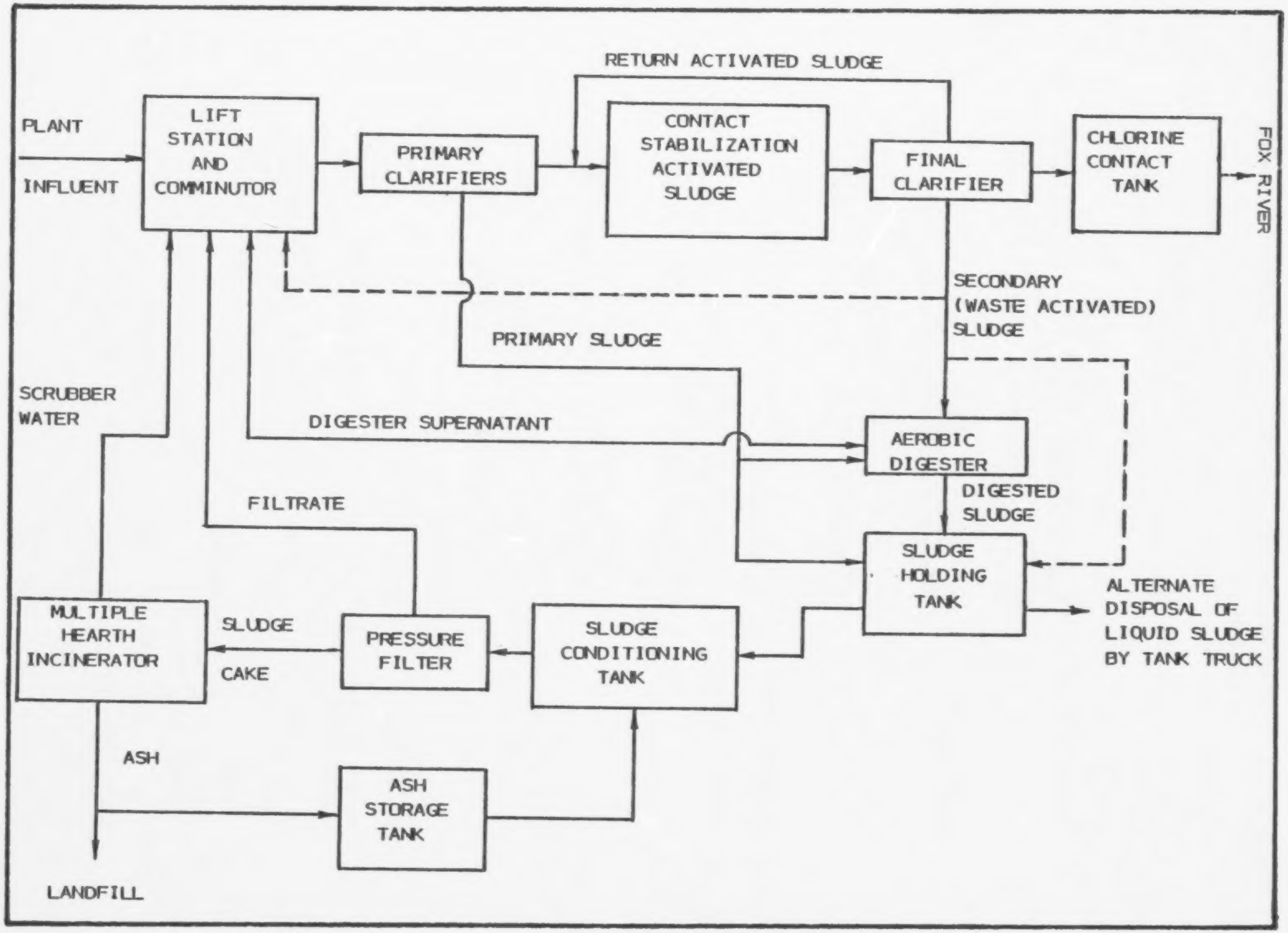

PLANT
INFLUENT
LIFT STATION AND COMMINUTOR
PRIMARY CLARIFIERS
RETURN ACTIVATED SLUDGE
CONTACT STABILIZATION ACTIVATED SLUDGE
FINAL CLARIFIER
CHLORINE CONTACT TANK
FOX RIVER
SECONDARY (WASTE ACTIVATED) SLUDGE
PRIMARY SLUDGE
SCRUBBER WATER
DIGESTER SUPERNATANT
AEROBIC DIGESTER
DIGESTED SLUDGE
FILTRATE
SLUDGE HOLDING TANK
ALTERNATE DISPOSAL OF LIQUID SLUDGE BY TANK TRUCK
MULTIPLE HEARTH INCINERATOR
SLUDGE
CAKE
PRESSURE FILTER
SLUDGE CONDITIONING TANK
ASH
ASH STORAGE TANK
LANDFILL

Figure C-17. Brookfield, Wisconsin, wastewater treatment plant flow diagram.

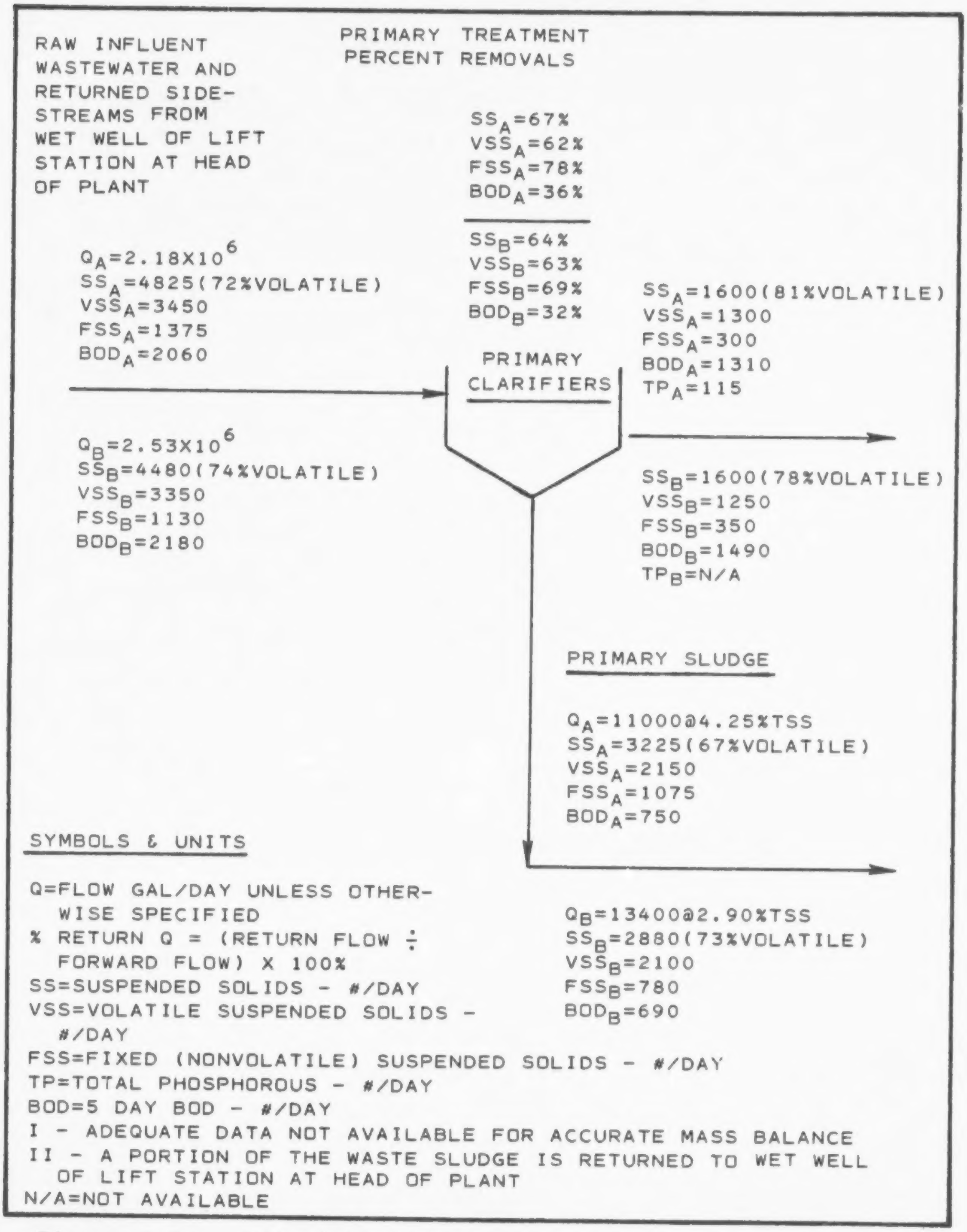
TABLE C-24. GENERAL PLANT DESCRIPTION SUMMARY, BROOKFIELD, WISCONSIN

Unit	No. and Description
Comminutor	
Lift Station	
Primary Clarifiers	2 at 16.80 m (55 ft) dia. x 2.44 m (8 ft) side wall depth (SWD); sludge collectors - 2 Walker Process Equipment type RSP with 2 collector arms; motors - 2 at 3/4 hp, 3Ø, 60 hz, 480 v.
Primary Sludge Helithickener	2 at 45.70 cm (18 in) dia. x 45.70 cm (18 in) pitch x 5.48 m (18 ft) length; Walker Process Equipment Helithickener Cross Collectors, 10-15 rpm; hopper dimensions - 2 at 21.3 m (70 ft) length x 0.762 m (2.5 ft) width x 1.680 m (5.5 ft) height; collector arm motors - 2 at 3/4 hp, 3Ø, 60 hz, 480 v.; helithickener motors - 2 at 1/2 hp, 3Ø, 60 hz, 480 z.
Primary Sludge Pumps	2 at 0.3 m^3/min (80 gal/min); 5 hp, 3Ø, 60 hz, 40 v.
Aeration Basins	3 at 28.30 m (93 ft) length x 9.140 m (30 ft) width x 4.570 (15 ft) height; air compressors - 3 at 2,190 scfm; motors - 3 squirrel cage induction motors at 100 hp, 3Ø, 60 hz, 480 v.
Air Diffusers	6 Walker Process Equipment EASEOUT header assemblies with 12 saddle-mounted MONOSPARJ diffusers (per basin) at 8 scfm/diffuser.
Final Clarifiers	2 at 22.90 m (75 ft) dia. x 3.660 m (12 ft) SWD; sludge collectors - 2 Walker Process Equipment type SWD equipped with surface skimmer and 2 collector arms with 7.62 to 15.2 (3 to 6 in) suction pipes; collector motors - 2 at 1/2 hp, 3Ø, 60 hz, 480 v.
Waste Activated Sludge Pump	1 at 0.3 m^3/min (80 gal/min); 5 hp, 3Ø, 60 hz, 480 v.
Return Activated Sludge Pumps	2 at 5,680ℓ/min (1,500 gal/min); 10 hp, 3Ø, 60 hz, 230 v.
Chlorine Contact Tank	1 at 12.20 m (40 ft) dia. x 3,660 m (12 ft) SWD
Lagoon	1 at 1,350 m^2 (14,500 ft^2) x 2.130 m (7 ft) depth
Aerobic Digester	1 at 11.0 m (36 ft) x 11.0 m (36 ft) x 7.62 m (25 ft) SWD with Walker Process Unit Equipment Rollaer diffusers
Sludge Processing Holding Tank	1 at 4.270 m (14 ft) x 4.270 m (14 ft) x 2.130 m (7 ft) SWD

(continued)

TABLE C-24 (continued)

Unit	No. and Description
Pressure Filter	1 plate and frame type Beloit-Passavant Series 5,200 - 46 chambers at 0.042 m^3 (1.5 ft^3) each with 112 m^3 (121 ft^2) total area; 241.0 kg (530 lb) dry solids/hr capacity; epoxy coated carbon steel construction filter plates; filter media-monofilament nylon with stainless steel 10 x 10 mesh backup wire; galvanized carbon steel drainage member.
Pressure Filter Appurtenances	1 mix tank at 4,390 m^2 (120 ft^3) with 2 agitators at 6 rpm; contact tank at 2.44 m (8 ft) dia. x 4.27 m (14 ft) length, 1,510 ℓ (4.850 gal) capacity; 3 rpm agitator powered by 3 hp TEFC motor; equalization tank at 1.680 m (5.5 ft) dia. x 2.680 m (8.8 ft) height, 1,610 ℓ (425 gal) capacity; filtrate weir tank; precoat tank at 13.30 m (3.5 ft) dia. x 23.6 m (8.8 ft) height; 2 sludge-transfer pumps at 189 ℓ (50 gal/min), 1 1/2 hp; pumps at 189 ℓ (50 gal)/min, 1 1/2 hp; precoat pump at 40 hp, 30, 60 hz, 460 v.
Incinerator	5 multiple hearths Nichols-Herreshoff sludge incinerators at 3.96 m (13 ft) outside depth x 5.33 m (17.5 ft) height; 242.0 kg (510 lb) dry solids/hr capacity fed at 1,230 kg (2,700 lb)/hr with ash at 1.5:1, moisture 50 percent, 27 percent volatile; auxilliary natural gas burners in hearths no. 3 and 5; operating temperature 667 to 871°C (1,200 to 1,600°F); induced draft exhaust gas fan at 661,000 cms^2/sec (1,400 cfm), 5 hp, 30, 60 hz, 480 v; ash screw conveyors.
Incinerator Exhaust Gas Scrubber	3-stage flooded tray type wet scrubber with quencher, scrubber, and cyclonic impingement separator; scrubber water pump at 25 hp, 30, 60 hz, 480 v.


RAW INFLUENT WASTEWATER AND RETURNED SIDE-STREAMS FROM WET WELL OF LIFT STATION AT HEAD OF PLANT
PRIMARY TREATMENT PERCENT REMOVALS
SS_A=67%
VSS_A=62%
FSS_A=78%
BOD_A=36%
SS_B=64%
VSS_B=63%
FSS_B=69%
BOD_B=32%
Q_A=2.18X10^6
SS_A=4825(72%VOLATILE)
VSS_A=3450
FSS_A=1375
BOD_A=2060
Q_B=2.53X10^6
SS_B=4480(74%VOLATILE)
VSS_B=3350
FSS_B=1130
BOD_B=2180
PRIMARY CLARIFIERS
SS_A=1600(81%VOLATILE)
VSS_A=1300
FSS_A=300
BOD_A=1310
TP_A=115
SS_B=1600(78%VOLATILE)
VSS_B=1250
FSS_B=350
BOD_B=1490
TP_B=N/A
PRIMARY SLUDGE
Q_A=11000@4.25%TSS
SS_A=3225(67%VOLATILE)
VSS_A=2150
FSS_A=1075
BOD_A=750
Q_B=13400@2.90%TSS
SS_B=2880(73%VOLATILE)
VSS_B=2100
FSS_B=780
BOD_B=690

SYMBOLS & UNITS

Q=FLOW GAL/DAY UNLESS OTHERWISE SPECIFIED

% RETURN Q = (RETURN FLOW ÷ FORWARD FLOW) X 100%

SS=SUSPENDED SOLIDS - #/DAY

VSS=VOLATILE SUSPENDED SOLIDS - #/DAY

FSS=FIXED (NONVOLATILE) SUSPENDED SOLIDS - #/DAY

TP=TOTAL PHOSPHOROUS - #/DAY

BOD=5 DAY BOD - #/DAY

I - ADEQUATE DATA NOT AVAILABLE FOR ACCURATE MASS BALANCE

II - A PORTION OF THE WASTE SLUDGE IS RETURNED TO WET WELL OF LIFT STATION AT HEAD OF PLANT

N/A=NOT AVAILABLE

Figure C-18. Materials balance for primary wastewater treatment operations at Brookfield, Wisconsin.

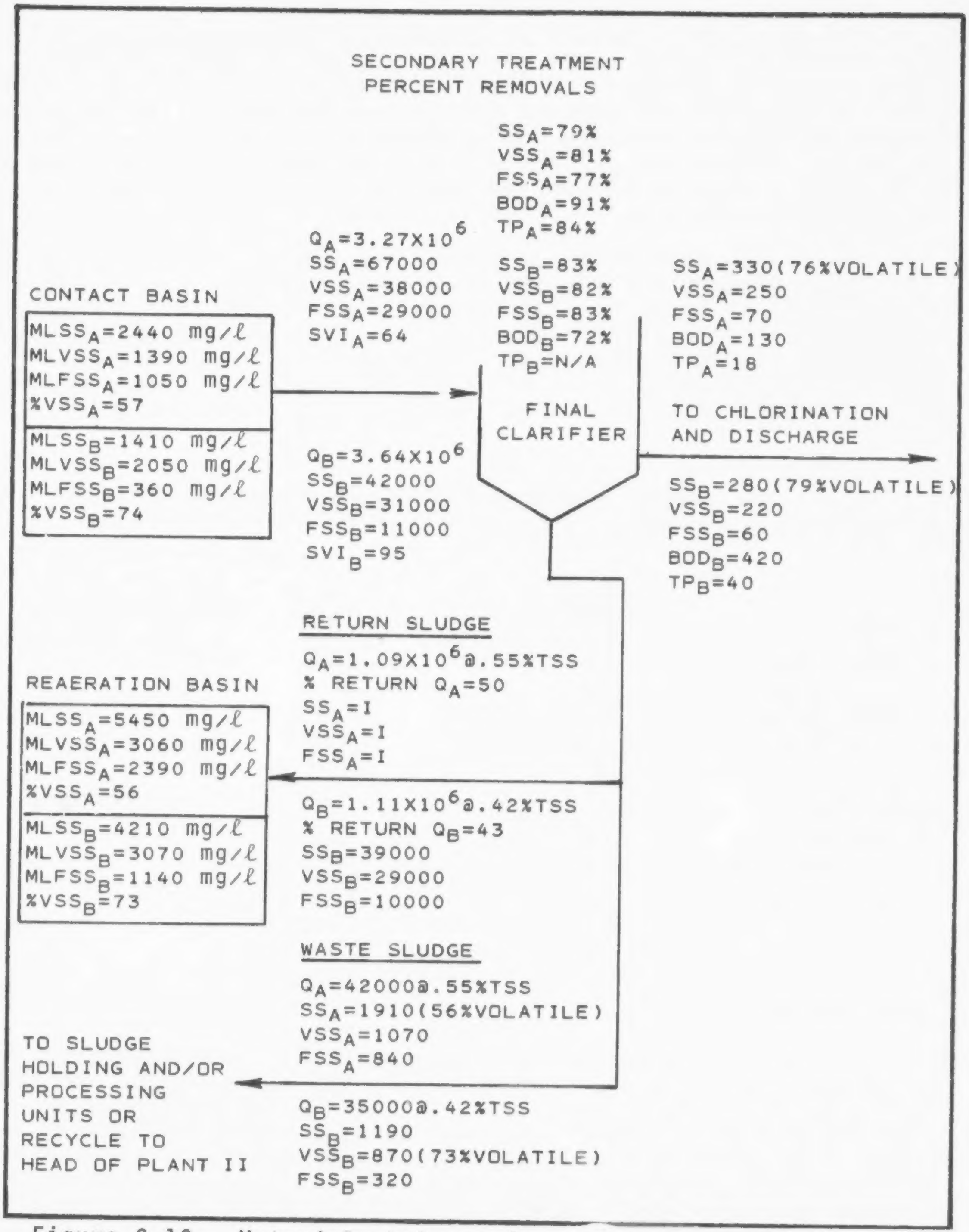

SECONDARY TREATMENT PERCENT REMOVALS
$SS_A=79\%$
$VSS_A=81\%$
$FSS_A=77\%$
$BOD_A=91\%$
$TP_A=84\%$
$SS_B=83\%$
$VSS_B=82\%$
$FSS_B=83\%$
$BOD_B=72\%$
TP_B=N/A
CONTACT BASIN
$MLSS_A=2440$ mg/ℓ
$MLVSS_A=1390$ mg/ℓ
$MLFSS_A=1050$ mg/ℓ
$\%VSS_A=57$
$MLSS_B=1410$ mg/ℓ
$MLVSS_B=2050$ mg/ℓ
$MLFSS_B=360$ mg/ℓ
$\%VSS_B=74$
$Q_A=3.27\times10^6$
$SS_A=67000$
$VSS_A=38000$
$FSS_A=29000$
$SVI_A=64$
$Q_B=3.64\times10^6$
$SS_B=42000$
$VSS_B=31000$
$FSS_B=11000$
$SVI_B=95$
FINAL CLARIFIER
$SS_A=330$(76%VOLATILE)
$VSS_A=250$
$FSS_A=70$
$BOD_A=130$
$TP_A=18$
TO CHLORINATION AND DISCHARGE
$SS_B=280$(79%VOLATILE)
$VSS_B=220$
$FSS_B=60$
$BOD_B=420$
$TP_B=40$
RETURN SLUDGE
$Q_A=1.09\times10^6$@.55%TSS
% RETURN $Q_A=50$
SS_A=I
VSS_A=I
FSS_A=I
$Q_B=1.11\times10^6$@.42%TSS
% RETURN $Q_B=43$
$SS_B=39000$
$VSS_B=29000$
$FSS_B=10000$
REAERATION BASIN
$MLSS_A=5450$ mg/ℓ
$MLVSS_A=3060$ mg/ℓ
$MLFSS_A=2390$ mg/ℓ
$\%VSS_A=56$
$MLSS_B=4210$ mg/ℓ
$MLVSS_B=3070$ mg/ℓ
$MLFSS_B=1140$ mg/ℓ
$\%VSS_B=73$
WASTE SLUDGE
$Q_A=42000$@.55%TSS
$SS_A=1910$(56%VOLATILE)
$VSS_A=1070$
$FSS_A=840$
$Q_B=35000$@.42%TSS
$SS_B=1190$
$VSS_B=870$(73%VOLATILE)
$FSS_B=320$
TO SLUDGE HOLDING AND/OR PROCESSING UNITS OR RECYCLE TO HEAD OF PLANT II

Figure C-19. Materials balance for secondary wastewater treatment operation at Brookfield, Wisconsin.

Primary Clarifiers--

Primary sedimentation of the raw influent wastewater and returned sidestreams takes place in the two primary clarifiers. Each clarifier is equipped with two sludge collector arms which continually move the settled solids towards the sludge hopper which contains a rotating "helithickener." Sludge in the helithickener hopper and collected clarifier surface scum are combined and pumped to the sludge holding and/or processing units one to three times daily, five days/week.

Since the initiation of phosphorus removal, the suspended solids loading on the primary clarifiers has increased by 27 percent or 202 kg SS/M^3 (445 lb SS/MG) of plant influent. However, since the average suspended solids concentration of the sludge increased from 2.9 percent to 3.5 percent SS, there was a net increase of 157 kg SS/M^3 (345 lb SS/MG) pumped to sludge processing. Further details of the impact on the primary clarifiers are shown in Table C-25.

TABLE C-25. CHANGES IN PRIMARY CLARIFIER PERFORMANCE AS THE RESULT OF CHEMICAL ADDITION FOR PHOSPHORUS REMOVAL, BROOKFIELD, WISCONSIN

Changes	SS kg/mil m^3 (lb/MG)	VSS kg/mil m^3 (lb/MG)	FSS* kg/mil m^3 (lb/MG)	BOD kg/mil m^3 (lb/MG)	Flow m^3/mil m^3 (gal/MG)
Primary Clarifier Influent Concentrations	+91,000 (+455†)	52,000 (+260)	+37,000 (+185)	+17,000 (+85)	NA#
Primary Clarifier Effluent Concentrations	+20,000 (+100)	+20,000 (+100)	No charge	+ 2,000 (+10)	NA
Primary Sludge Pumped	(+69,000) (+345)	32,000 (+160)	+37,000 (+185)	+15,000 (+75)	-50,000 (-250)

*FSS = non-volatile (fixed) suspended solids.

†+ indicates increase; - indicated decrease.

#NA = Not Applicable.

Aeration Basins--

Currently the activated sludge system is being operated as a contact stabilization process, using one of the three aeration basins as the contact basin, and one as the reaeration basin. (The third basin is operated as an aerated sludge storage tank apart from the activated sludge process). Prior to entering the contact basin, primary effluent is received by a preaeration chamber where it is mixed with flow from the reaeration basin.

Since the initiation of chemical addition for phosphorus removal, the average MLSS concentration in the contact basin has increased by nearly 75 percent, with the majority of this increase due to a 190 percent increase in the non-volatile (fixed) portion. As a result, the VSS fraction of MLSS decreased from 74 percent VSS prior to phosphorus removal to 57 percent VSS since phosphorus removal was initiated. Since there was virtually no change in the amount of nonvolatile SS entering the contact basin from the primary clarifier, it is assumed that the entire increase is due to the addition of phosphorus removal chemicals and subsequent reactions.

In the reaeration basin, the MLSS concentration has increased by nearly 30 percent since phosphorus removal was begun. The entire change is due to a 110 percent increase in the concentration of fixed SS in the mixed liquor (MLFSS). Consequently, the volatile fraction of the reaeration basin mixed liquor (MLVSS) decreased from 73 percent to 56 percent VSS.

Final Clarifiers--

The final clarifiers receive the mixed liquor effluent from the contact basin. Sludge is withdrawn from the bottoms of the clarifiers on a continuous basis and pumped as return sludge to the reaeration basin. As necessary, sludge is wasted to one of the sludge holding tanks, sludge processing units, or returned to the head of the plant. Roughly 25 percent of the waste sludge is returned directly to the head of the plant. Thus, approximately 75 percent is sent to one of the sludge holding or processing units. (The portion sent to the aeration basin not in use for secondary treatment is later returned to the head of the plant.)

Since phosphorus removal was begun, the settleability of the secondary sludge has improved. This was evidenced by a decrease in sludge volume index from an SVI of 99 prior to phosphorus removal to an SVI of 64 since phosphorus removal was begun.

After phosphorus removal, the pumping of return activated sludge from the final clarifier to the reaeration basin increased from 43 percent to 50 percent of the plant's forward flow. The amount of waste activated sludge pumped also increased by 5,430 m^3/mil m^3 (5,430 gal/MG) of plant influent, a 40 percent increase. At the same time, the return and waste activated sludge SS concentrations increased from 4,200 mg/ℓ SS to 5,500 mg/ℓ SS, causing an additional 48,600 kg SS/mil m^3 (405 lb SS/MG) of plant influent to be wasted. Table C-26 shows these impacts of chemical addition for phosphorus removal, as well as the impacts on final clarifier effluent.

TABLE C-26. IMPACTS OF CHEMICAL ADDITION FOR PHOSPHORUS REMOVAL ON FINAL CLARIFIERS, BROOKFIELD, WISCONSIN

Changes	Flow m^3/mil m^3 (gal/MG)	SS kg/mil m^3 (lg/MG)	VSS kg/mil m^3 (lb/MG)	FSS kg/mil m^3 (lb/MG)	BOD
Final Clarifier Effluent Concentrations	NA[†]	+ 4,800 (+40)	+ 3,000 (+25)	+ 1,800 (+15)	-13,200 (-110)
Waste Activated Sludge Pumped	+5,430	+48,600 (+405)	+17,400 (+145)	+31,000 (260)	NA
Returned Activated Sludge Pumped*	$+0.061 \times 10^6$	---	---	---	NA

*Adequate data not available for accurate mass balance.

[†] NA = Not Applicable

Sludge Treatment and Disposal Operations

After sludge has been removed from the primary or final clarifiers, it can be stored prior to sludge processing in three locations as follows:

- Aeration basin (not in use for secondary treatment)
- Aerobic digester
- Sludge processing holding tank.

Unfortunately, it was not possible to thoroughly assess the impact of chemical addition for phosphorus removal on each of the individual sludge storage units. This was due to the wide variability of sludge storage and disposition operations and the shortness and discontinuous nature of the periods available for study in comparison to the retention times of the individual sludge holding units. An additional problem in evaluating impacts has been the necessity to use suspended solids concentration for constructing mass balances around clarifiers and the activated sludge process, but to use total solids concentrations throughout the discussion of sludge processing.

The remainder of this section discusses each of the sludge storage units, sludge volume reduction units, and disposal operations was shown previously in Figure C-17. A diagram of sludge treatment and disposal facilities is shown in Figure C-20. Flow diagrams showing quantities, and characteristics of sludge and conditioning chemicals fed to the pressure filter and incinerator during the periods before and after phosphorus removal, are shown in Figures C-21 and C-22.

Aeration Basin (not in use for secondary treatment)--

At present, only two of the three aeration basins are used as part of the activated sludge system. Instead, the third is used to provide additional aerated sludge storage capacity. An

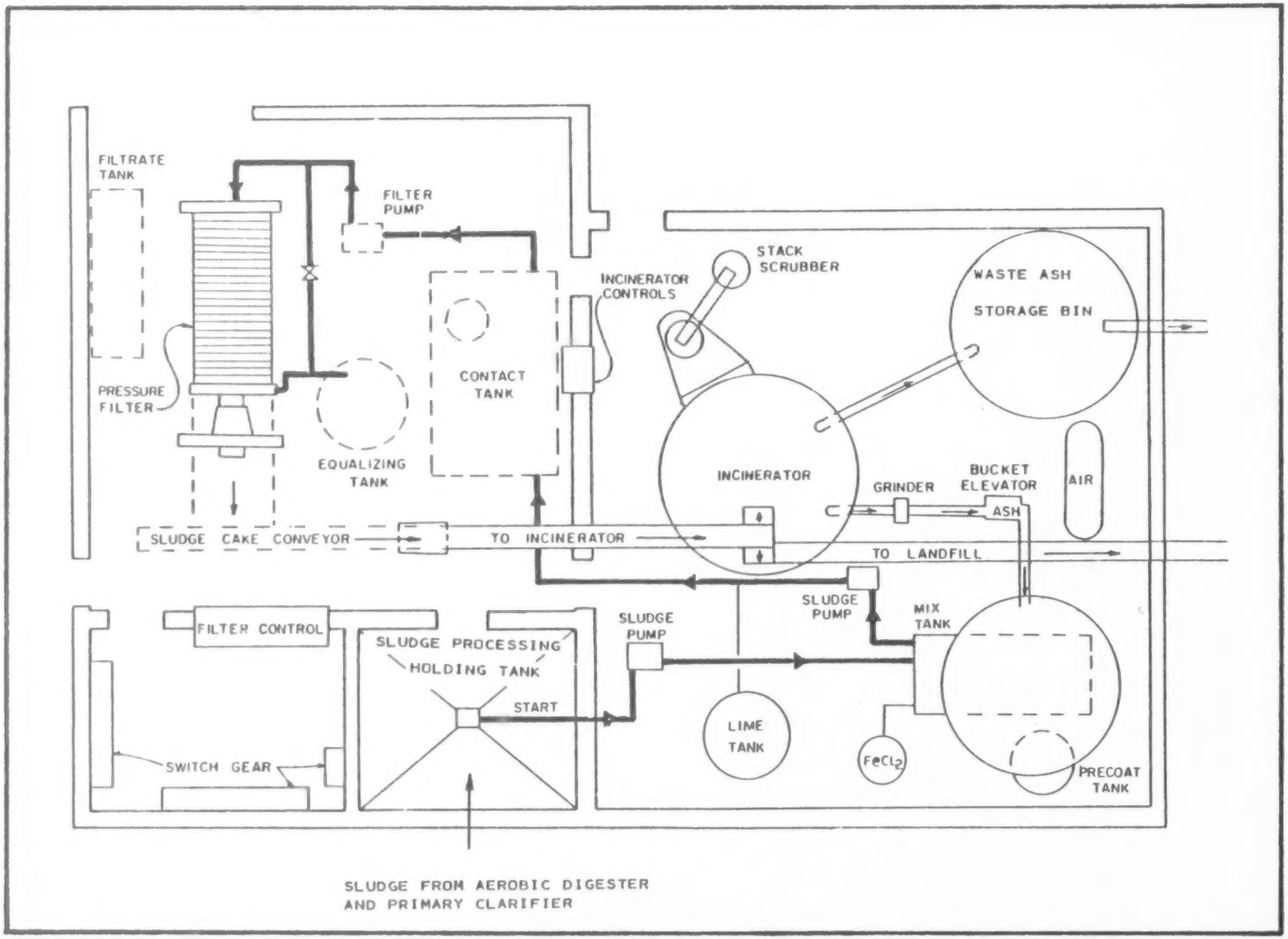

FILTRATE TANK
PRESSURE FILTER
FILTER PUMP
EQUALIZING TANK
CONTACT TANK
INCINERATOR CONTROLS
STACK SCRUBBER
WASTE ASH STORAGE BIN
INCINERATOR
GRINDER
BUCKET ELEVATOR
ASH
AIR
SLUDGE CAKE CONVEYOR
TO INCINERATOR
TO LANDFILL
SLUDGE PUMP
MIX TANK
SLUDGE PUMP
FILTER CONTROL
SLUDGE PROCESSING
HOLDING TANK
START
LIME TANK
$FeCl_2$
PRECOAT TANK
SWITCH GEAR
SLUDGE FROM AEROBIC DIGESTER AND PRIMARY CLARIFIER

Figure C-20. Brookfield, Wisconsin, pressure filtration and incineration facilities.

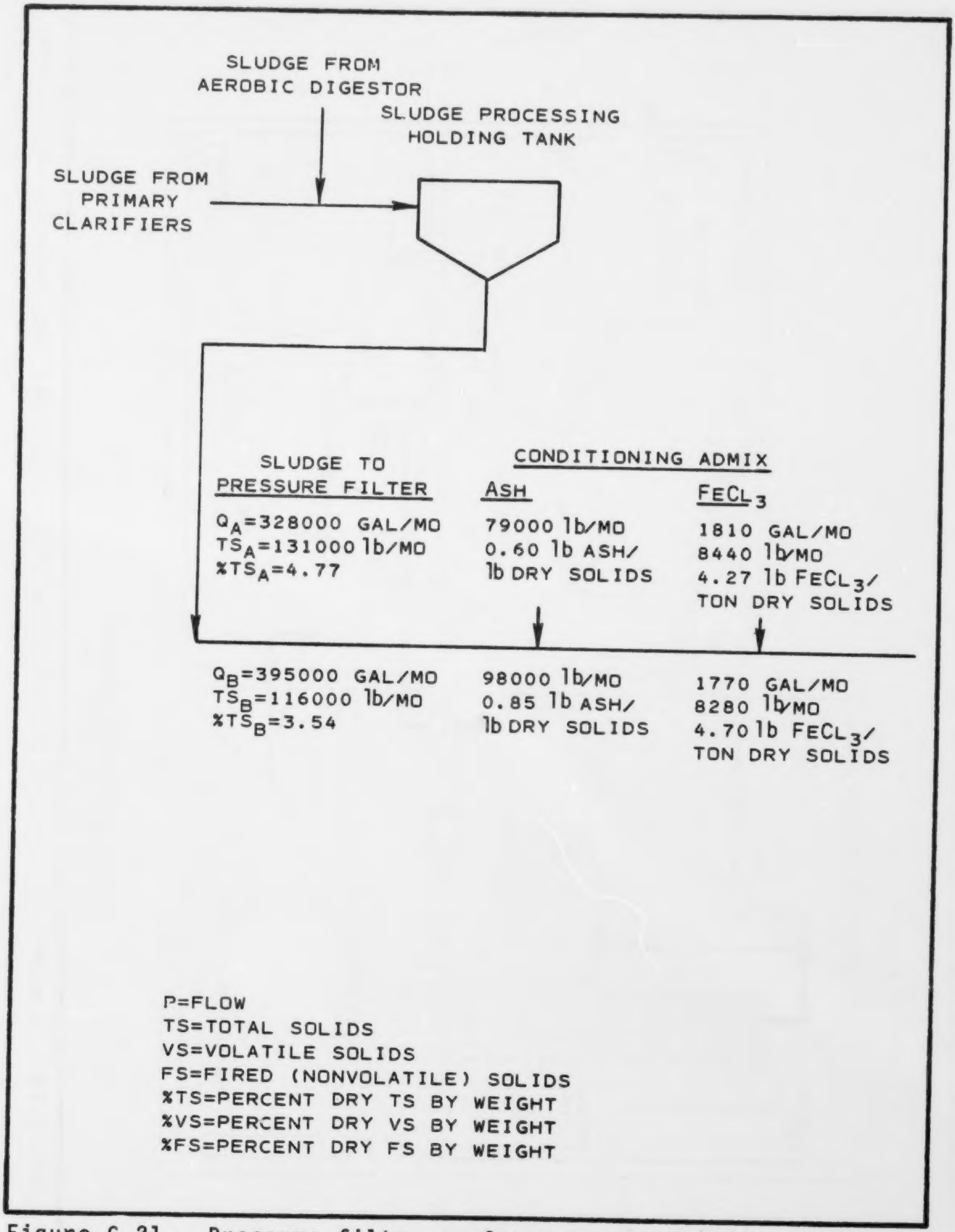

SLUDGE FROM AEROBIC DIGESTOR
SLUDGE PROCESSING HOLDING TANK
SLUDGE FROM PRIMARY CLARIFIERS
SLUDGE TO PRESSURE FILTER
CONDITIONING ADMIX
ASH
$FeCl_3$
Q_A=328000 GAL/MO
TS_A=131000 lb/MO
$\%TS_A$=4.77
79000 lb/MO
0.60 lb ASH/lb DRY SOLIDS
1810 GAL/MO
8440 lb/MO
4.27 lb $FeCl_3$/TON DRY SOLIDS
Q_B=395000 GAL/MO
TS_B=116000 lb/MO
$\%TS_B$=3.54
98000 lb/MO
0.85 lb ASH/lb DRY SOLIDS
1770 GAL/MO
8280 lb/MO
4.70 lb $FeCl_3$/TON DRY SOLIDS
P=FLOW
TS=TOTAL SOLIDS
VS=VOLATILE SOLIDS
FS=FIRED (NONVOLATILE) SOLIDS
%TS=PERCENT DRY TS BY WEIGHT
%VS=PERCENT DRY VS BY WEIGHT
%FS=PERCENT DRY FS BY WEIGHT

Figure C-21. Pressure filter performance, Brookfield, Wisconsin.

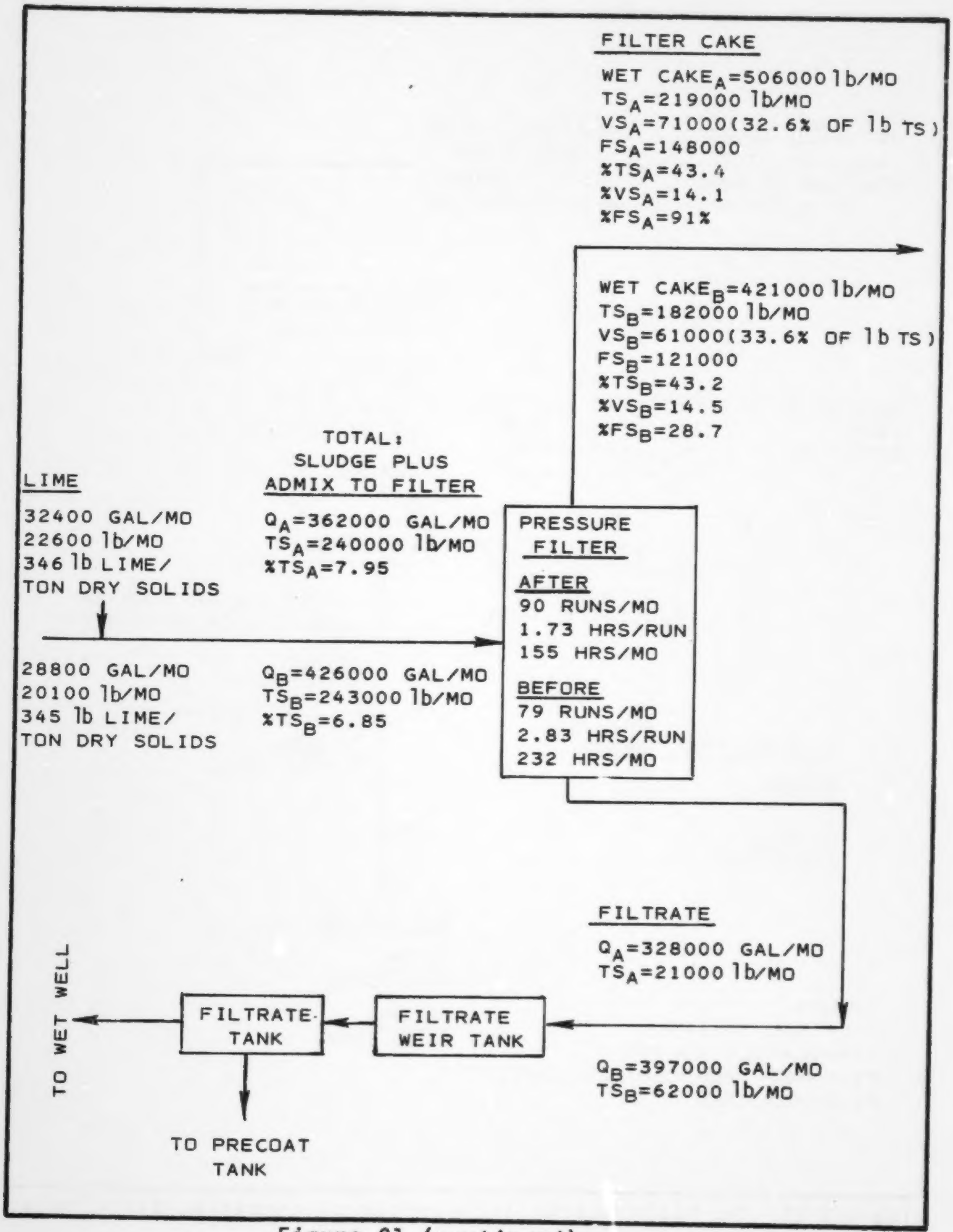

FILTER CAKE
WET CAKE$_A$=506000 lb/MO
TS$_A$=219000 lb/MO
VS$_A$=71000(32.6% OF lb TS)
FS$_A$=148000
%TS$_A$=43.4
%VS$_A$=14.1
%FS$_A$=91%
WET CAKE$_B$=421000 lb/MO
TS$_B$=182000 lb/MO
VS$_B$=61000(33.6% OF lb TS)
FS$_B$=121000
%TS$_B$=43.2
%VS$_B$=14.5
%FS$_B$=28.7
LIME
32400 GAL/MO
22600 lb/MO
346 lb LIME/
TON DRY SOLIDS
28800 GAL/MO
20100 lb/MO
345 lb LIME/
TON DRY SOLIDS
TOTAL:
SLUDGE PLUS
ADMIX TO FILTER
Q$_A$=362000 GAL/MO
TS$_A$=240000 lb/MO
%TS$_A$=7.95
Q$_B$=426000 GAL/MO
TS$_B$=243000 lb/MO
%TS$_B$=6.85
PRESSURE
FILTER
AFTER
90 RUNS/MO
1.73 HRS/RUN
155 HRS/MO
BEFORE
79 RUNS/MO
2.83 HRS/RUN
232 HRS/MO
FILTRATE
Q$_A$=328000 GAL/MO
TS$_A$=21000 lb/MO
Q$_B$=397000 GAL/MO
TS$_B$=62000 lb/MO
FILTRATE WEIR TANK
FILTRATE TANK
TO WET WELL
TO PRECOAT TANK

Figure 21 (continued)

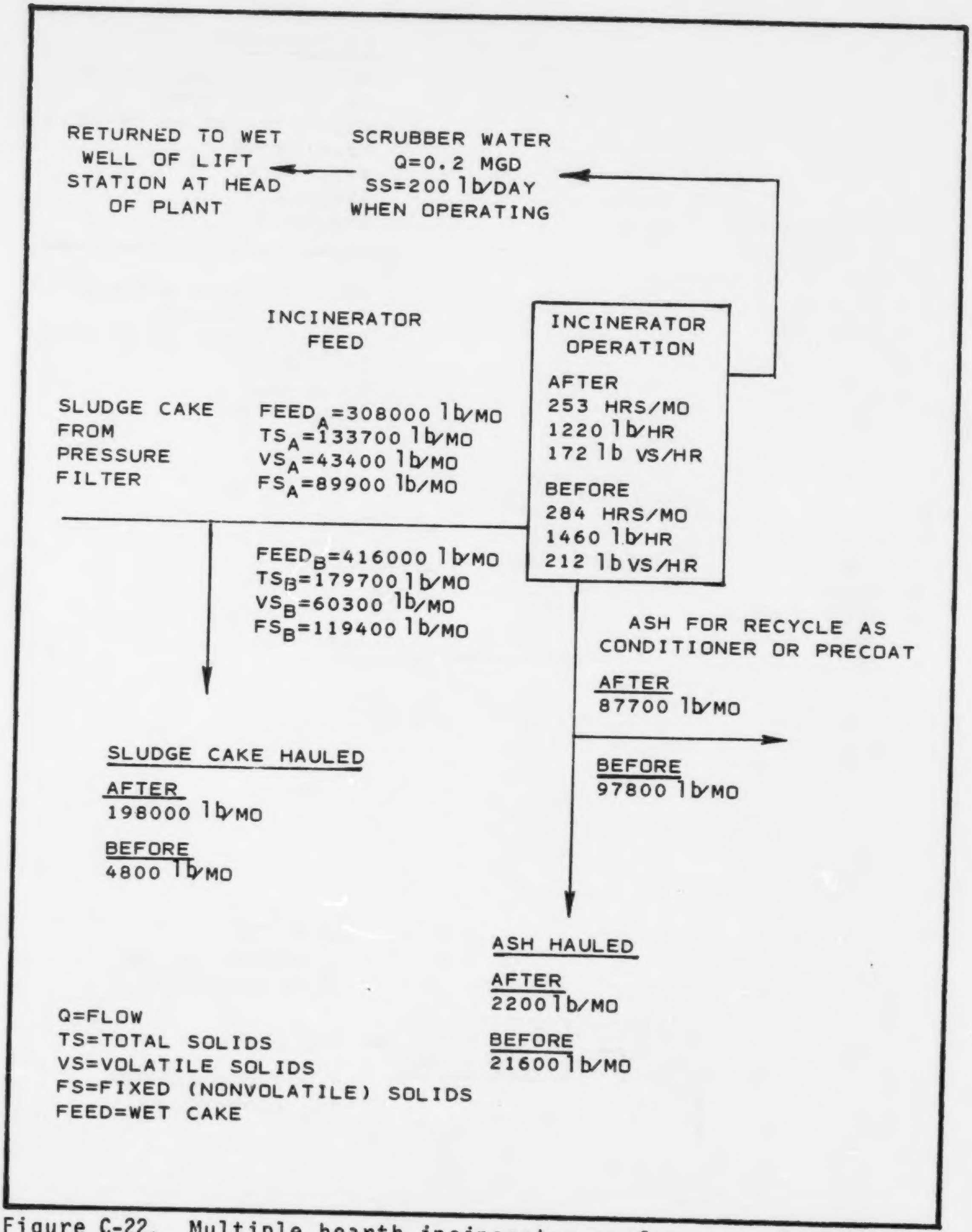

RETURNED TO WET WELL OF LIFT STATION AT HEAD OF PLANT
SCRUBBER WATER
Q=0.2 MGD
SS=200 lb/DAY
WHEN OPERATING
INCINERATOR FEED
SLUDGE CAKE FROM PRESSURE FILTER
FEED$_A$=308000 lb/MO
TS$_A$=133700 lb/MO
VS$_A$=43400 lb/MO
FS$_A$=89900 lb/MO
INCINERATOR OPERATION
AFTER
253 HRS/MO
1220 lb/HR
172 lb VS/HR
BEFORE
284 HRS/MO
1460 lb/HR
212 lb VS/HR
FEED$_B$=416000 lb/MO
TS$_B$=179700 lb/MO
VS$_B$=60300 lb/MO
FS$_B$=119400 lb/MO
ASH FOR RECYCLE AS CONDITIONER OR PRECOAT
AFTER
87700 lb/MO
BEFORE
97800 lb/MO
SLUDGE CAKE HAULED
AFTER
198000 lb/MO
BEFORE
4800 lb/MO
ASH HAULED
AFTER
2200 lb/MO
BEFORE
21600 lb/MO
Q=FLOW
TS=TOTAL SOLIDS
VS=VOLATILE SOLIDS
FS=FIXED (NONVOLATILE) SOLIDS
FEED=WET CAKE

Figure C-22. Multiple hearth incinerator performance, Brookfield, Wisconsin.

estimated 60 percent of the plant's waste activated sludge and 5 percent of the primary sludge enters this unit by pumping as necessary from the bottoms of the primary or final clarifiers. Since the aeration basin is not equipped for the collection and pumping of concentrated sludge, solids contained in the aeration basin are pumped from the aeration basin without being allowed to settle. This sludge is then sent to the wet well of the lift station at the head of the plant. Subsequently, it is pumped to the primary clarifiers along with the raw influent wastewater and other returned sidestreams. There, the combined raw and returned solids are settled and concentrated, after which they are generally sent to one of the sludge holding units other than the aeration basin.

Aerobic Digester--

An estimated 80 percent of the sludge routed to the aerobic digester is from the primary clarifiers while 20 percent is from the final clarifiers. The aerobic digester supernatant is decanted periodically to the wet well at the head of the plant, after the aeration equipment has been turned off and the solids have settled to the bottom of the digester. The sludge is pumped from the bottom of the digester to the sludge processing holding tank when the aeration equipment has been turned off. Since phosphorus removal was initiated, the aerobically digested sludge has averaged 1.1 percent TS. Before phosphorus removal began, the sludge contained 1.5 percent TS.

Sludge Processing Holding Tank--

An estimated 85 percent of the sludge routed to this unit is from the primary clarifiers and 15 percent from the aerobic digester. Occasionally, supernatant is decanted from this tank and returned to the wet well at the head of the plant. Sludge from the sludge processing holding tank is generally fed to the pressure filter (although liquid sludge can be pumped into a tank truck and hauled from the plant for spreading on croplands). Figure C-21 presents data on the average characteristics of the sludge from the sludge processing holding tank.

Pressure Filter--

The plate and frame pressure filter is generally operated 4 days/wk, 16 hr/day, 45 wk/yr. (To facilitate data comparisons, sludge quantities and flow rates are expressed as daily amounts per 7-day week.) A filter run begins by pumping sludge from the sludge processing holding tank (sludge supply tank) to the mix tank where it is blended with admix materials into a uniform slurry. The admix material consists of incinerator ash, ferric chloride solution of 570 kg $FeCl_3/m^3$ (4.75 lb $FeCl_3$/gal) of solution, and hydrated lime diluted to an 8 percent solution.

Next, the sludge and conditioning admix slurry are pumped from the mix tank to the contact tank. This tank is provided to allow sufficient residence time for conditioning reactions to take place prior to feed to the filter. The tank is equipped with an agitator blade which turns at 3 rpm to prevent the suspension from clinging to the tank walls. An equalizing tank receives the slurry from the contact tank and holds it until pumping from the tank to the filter begins.

Prior to filtration, precoat consisting of filtrate and incinerator ash fed from the precoat tank at 100 psi to cause is a uniform precoat slurry to flow through the filter. Following filter precoating, sludge and admix from the equalizing tank are pumped to the filter, initially at 379 ℓ (100 gal/min) and 225 psi pressure to preclude the loss of precoat from the filter media. When the tank has delivered approximately one half of the conditioned sludge to the filter, an air supply valve is opened to complete the emptying of the tank at a reduced but constant pressure of 100 psi. Throughout the filter run, the filter feed pump automatically decreases output as the pumping head increases, and eventually stalls out against a terminal pressure of 225 psi (which coincides with zero flow through the filter).

Upon completion of the filtration cycle, sludge cake is discharged from the filter and dropped into a bunker where it is stored. A portion of the filtrate from the pressure filter operation is returned to the wet well at the head of the plant while the remainder of the filtrate is returned to the precoat tank. Sludge cake is removed from the bunker by a drag flight conveyor, discharged onto an inclined conveyor, and then fed to the incinerator. Alternatively, if may be discharged to a bypass conveyor which discharges the sludge cake to a truck outside the building.

Data describing the operation of the pressure filter during the period before and after phosphorus removal was initiated is shown in Table C-27. After phosphorus removal, there was a slight decrease in the volume of sludge fed from the sludge processing tank to the chemical mix tank each month which is equal to a decrease of 185 m^3/mil m^3 (185 gal/MG) of plant influent. However, the total solids in the sludge holding tank increased from 3.54 percent TS before phosphorus removal to 4.77 percent TS after phosphorus removal. As a result the mass of total solids fed to the chemical mix tank from the sludge processing holding tank has increased by 30 percent, or 56,405 kg/mil m^3 (470 lbs/MG) of plant influent.

TABLE C-27. PRESSURE FILTER PERFORMANCE, BROOKFIELD, WISCONSIN

	Pressure Filter Input	
	After	Before
m^3 (gal) feed/run	13.7 (3,640)	18.9 (5,000)
m^3 (gal) feed and admix run	15.2 (4.020)	20.4 (5,380)
kg (lb) TS feed/run	661.0 (1,455)	667.0 (1,470)
kg (lb) TS feed and admix run	1,200 (2,670)	1,400 (3,075)
kg (lb) TS feed/hr	384.0 (845)	227 (500)
kg (lb) TS feed and admix/hr	704.0 (1,550)	477.0 (1,050)
	Pressure Filter Output	
	After	Before
kg (lb) wet cake/run	2,550 (5,620)	2,420 (5,330)
kg (lb) wet cake/hr	1,480 (3,265)	824 (1,815)
kg (lb) TS in cake/run	1,100 (2,430)	1,044 (2,300)
kg (lb) TS in cake/hr	638 (1,405)	370 (815)
kg (lb) TS in filtrate/run	107 (235)	356 (785)
kg (lb) TS in filtrate/hr	61.3 (135)	125 (275)
% feed and admix TS recovered in cake	90	75

When the sludge gets to the chemical mix tank, incinerator ash, $FeCl_3$, and lime are added, again changing the quantity and solids concentration of the sludge before it gets to the filter.

The dosage of ash was lower by about 0.1140 kg ash/kg of dry solids (0.25 lb ash/lb of dry solids) after phosphorus removal. The average dosage decreased from 0.386 to 0.272 kg ash/kg of dry solids (0.85 to 0.60 lb ash/lb of dry solids). Similarly, the dosage of ferric chloride decreased from 2.35 kg $FeCl_3$/t (4.70 lb $FeCl_3$/ton) of dry solids to 2.14 kg $FeCl_3$/t (4.27 lb $FeCl_3$/ton) of dry solids after phosphorus removal. The dosage of lime remained constant at approximately 173.0 kg lime/t (345 lb lime/ton) of dry solids. The net result was that the volume of sludge plus admix delivered to the filter from the mix tank was lower by 80 m^3/mil m^3 (80 gal/MG) of plant influent after phosphorus removal. And there was a total solids increase from 6.85 percent TS to 7.95 percent TS in the sludge fed to the filter. This is equivalent to an increase in the amount of pressure filter feed of 55,200 kg TS/mil m^3 (460 lb TS/MG) of plant influent, or an increase of 15 percent.

Filter cake characteristics showed only minor variations between the periods before and after phosphorus removal. Specifically, the TS concentration of the cake increased from 43.2 to 43.4 percent. The VS fraction of cake TS decreased somewhat from 33.6 percent of TS to 32.6 percent of TS. Table C-27 compares pressure filter performance before and after phosphorus removal.

Since the initiation of chemical addition for phosphorus removal, pressure filter performance has improved significantly. The sludge mass fed per hour (not including admix materials) has increased from 227 to 384 kg TS/hr (500 to 845 lb TS/hr), and the mass of feed plus admix/hr increased from 477 to 704 kg/hr (1,050 to 1,550 lb/hr). The length of the average pressure filter run decreased by 40 percent from 2.83 hrs/run to 1.73 hrs/run. In addition, the percentage of the solids fed to the filter which are recovered in the cake increased from 75 percent to 90 percent. The combined effect was a production increase of 658 kg wet cake/hr (1,450 lb wet cake/hr) and 268 kg TS/hr (590 lb TS/hr).

Incinerator--

During periods of incinerator operation, the incinerator is run 24 hrs/day, 4 days/wk with an additional half day for incinerator start-up, and another half day for cool-down. The cake elevating conveyor raises pulverized filter cake to the top of the incinerator, where the cake enters the first hearth. After the cake has been fully incinerated and reaches the bottom hearth, the ash is discharged into a screw conveyor which in turn discharges to an ash grinder, followed by a bucket elevator which ultimately discharges into the ash storage bin. Ash stored in the bin is used for precoating the pressure filter or conditioning the sludge prior to pressure filtration. Excess ash can be discharged from the ash bin into a truck for disposal on-site followed by subsequent transfer to an off-site landfill.

Exhaust from the incinerator is either vented to the air pollution control system or returned to the lower hearths through the warm air duct for preheating. The air pollution control equipment consists of a 3-stage, flooded-tray type wet scrubber. The first stage consists of a hot gas quencher where water is sprayed into the gases until they are cooled. The second stage is the scrubber where water is sprayed to entrain particulate matter. Finally, the gases are exhausted through the third stage which consists of a cyclonic impingement separator followed by the exhaust stack. Scrubber waters pass through the scrubber once, after which they are discharged to the wet well at the head of the plant.

The operating temperatures recorded at the hearths and the scrubber system are:

- Hearth 1 - 427°C (800°F)
- Hearth 2 - 867°C (900°F)
- Hearth 3 - 843°C (1,550°F)
- Hearth 4 - 704°C (1,300°F)
- Hearth 5 - 468°C (600°F)
- Cooling Air Outlet - 38°C (100°F)
- Scrubber Inlet - 93°C (200°F)
- Scrubber Outlet - 248°C (280°F)

Incinerator fuel consumption is approximately 18.5 to 19.6 m^3/hr (660 to 700 ft^3/hr) of operation (natural gas).

Figure C-22 shows performance data for the multiple hearth incinerator. From this data and the information presented in the general plant description summary (Table C-24), the incinerator can be seen to be presently operating at an average feed rate which is approximately one-half of the design capacity. Although not attributable to phosphorus removal, there has been a shift in the method of final disposal used. Before phosphorus removal an average of 1.8 t/mo (2 tons/mo) wet pressure filter cake were hauled from the plant rather than incinerated. More recently, nearly 90.7 t/mo (100 tons/mo) of wet cake are being hauled. The plant manager apparently has found it cost effective to keep the incinerator down for a week rather than operate at less than capacity. Currently, consideration is being given to hauling all the wet sludge cake produced as a cost effective alternative to incineration.

Since the beginning of phosphorus removal, incinerator feed rates have declined from 663.0 to 554.0 kg wet cake/hr (1,460 to 1,220 lb wet cake/hr) of operation. This has reduced the feed rate of volatile solids from 96.3 to 78.1 kg VS/hr (212 to 172 lb VS/hr) of operation. As a result, an additional 14 m^3 (500 ft^3) of natural gas are required per hour of operation.

Prior to the initiation of phosphorus removal, approximately 9.98 t (11 tons) of incinerator ash were hauled from the plant each month, while an average of only 0.907 t (1 ton) of ash was hauled each month after phosphorus removal was initiated. The remainder of the ash generated by the incinerator is recycled as admix for pressure filter sludge conditioning or pressure filter precoat. During periods of incinerator operation, approximately 757 m^3/day (0.2 mgd) of scrubber water containing approximately 90.8 kg (200 lb) SS/day are generated and returned to the head of the plant.

Sludge Treatment and Disposal Costs

Operation and Maintenance Costs--

With the initiation of chemical addition for phosphorus removal, the cost of sludge treatment and disposal has decreased by approximately \$1.60/t (\$1.33/ton) of dry solids filtered, as shown in Table C-28. This reduction was due to decreases in the amounts of chemical conditioners and electricity used by the plate and frame pressure filter. These decreases were partially offset by an increase in the amount of auxiliary fuel used by the incinerator as the result of decreased incinerator volatile solids feed rates.

The mass of solids filtered increased from 182 to 237 t/mil m^3 (0.76 to 0.99 tons/MG) of plant influent with phosphorus removal. Theoretically, sludge treatment costs should have increased by 28 percent going from \$179,000.00 to \$229,000.00 (\$47.30 to \$60.50/MG) of plant influent. However, increased wet sludge cake hauling (at little or no charge) and reduced periods of incinerator operation prevented any actual increase in the cost of sludge treatment per MG of plant influent.

Capital Costs--

The total plant cost in 1973 was \$3.6 million. The plant was financed by 15 year general obligation bonds at 4.37 percent interest, maturing in 1986. The pressure filter equipment and piping cost \$385,000.00. The cost of the multiple hearth incinerator was \$336,000.00.

Summary and Conclusions

Since the initiation of chemical addition for phosphorus removal, the Brookfield wastewater treatment plant has shown an increase in the average solids loadings to each wastewater treatment and sludge handling unit process. The only exception to this observation was a decrease in total amount of sludge solids sent to the incinerator, which was due to increased wet sludge cake hauling after pressure filtration. There has been an increase of approximately 0.056 kg TS/m^3 (470 lb TS/MG) of plant influent pressure filtered as the result of phosphorus removal. Fortunately, due to the improved dewatering performance of the pressure filter, presumably attributable to phosphorus removal chemical addition, conditioning admix dosages of ash and ferric chloride were reduced. In addition, the total solids recovery by the pressure filter also improved significantly, improving filtrate quality.

Since this plant is operating well below capacity, the additional phosphorus laden chemical sludge at this plant has caused no operational problems. Furthermore, since wet sludge cake can

TABLE C-28. PRESSURE FILTER AND INCINERATOR OPERATIONAL COSTS PER t (ton) DRY SOLIDS, BROOKFIELD, WISCONSIN

Item	Unit Cost $	Units Used per t (ton) dry solids		Costs per t (ton) Dry Solids		
		After	Before	After	Before	Change
$FeCl_3$*	83.20/m^3 (0.315/gal)	0.115 (27.6)	0.13 (30.5)	9.58 (8.69)	10.59 (9.61)	-1.01 (-0.92)
Lime	0.067/kg (0.0305/lb)	173 (346)	172 (345)	11.63 (10.55)	11.60 (10.52)	+0.03 (+0.03)
Natural† Gas	0.064/m^3 (0.001786/ft^3)	(213) 6,880	(204) 6,570	13.55 (12.29)	12.93 (11.73)	+0.62 (+0.56)
Electricity#	0.04/kwh	240	260	10.58 (9.60)	11.47 (10.40)	-0.88 -0.80
Labor**	6.00/hr	3.33	3.33	22.05 (20.00)	22.05 (20.00)	No Change
(O&M) TOTAL				67.39 (61.13)	68.64 (62.26)	-1.24 -1.13

*During 1976, the plant installed ferric chloride bulk storage facilities reducing ferric chloride unit costs \$206/m^3 (\$0.780/gal), \$1,190/m^3 (\$0.315/gal). The unit cost for bulk purchases has been used for the purposes of cost comparisons.

†Includes incinerator warm-up.

#Estimated decrease due to shorter length of filter press runs.

**Assumes no sludge cake hauled from plant.

be hauled from the plant at little or no charge, increased quantities of more easily dewaterable sludge have actually decreased the cost of sludge handling per ton of dry solids.

CASE STUDY I: MIDLAND, MICHIGAN

Introduction

In 1960, when the Midland plant was designed as a two-stage, high-rate trickling filter plant, various alternative means of sludge handling were considered. The one chosen was vacuum filtration of the undigested sludge with disposal in a sanitary landfill. In 1970, it became necessary to upgrade the Midland plant to provide for phosphorus removal and a higher level of BOD and SS removal. Facilities were added for chemical additions of either ferric chloride or alum plus polymer, and rapid sand filters were added to filter the effluent prior to discharge. These changes meant that it would be necessary to handle additional sludge solids from the chemical addition and the tertiary filtration both, and this was considered in the design of the sludge handling facilities.

Besides the increased mass of sludge to be disposed of, two other factors were considered in the design of the new sludge handling facilities. One was the new requirement by the Michigan Department of Natural Resources that sludge stabilization and dewatering to at least 50 percent dry solids be performed if the plant was to continue sanitary landfill disposal. Previously, raw sludge at a concentration of about 25 percent TS had been landfilled. The other factor that was considered was cost.

Of the various alternatives considered for the solids handling facilities, the addition of thermal conditioning with continued utilization of the existing vacuum filters and sanitary landfill disposal appeared the least costly. Other alternatives considered included anaerobic digestion and farmland disposal of liquid sludge or sludge cake; incineration of sludge cake; and replacing the vacuum filters with a filter press. Thermal conditioning was the only method which could provide both sludge stabilization and a 50 percent TS filter cake. It thus enabled continued use of the sanitary landfill as well as the greatest reduction in the volume of sludge for disposal. Thermal conditioning also appeared to require less operational labor than any of the other alternatives.

When the thermal conditioning option was selected in 1970, the energy shortage was not a factor considered. At the present time, the alternatives are being re-evaluated by Midland due to the increased awareness of the energy shortage. This case study will investigate the impact of ferric chloride and polymer addition ahead of the primary clarifiers on the cost-effectiveness of the present system. Other plants considering the use of thermal conditioning and vacuum filtration can benefit from this discussion.

The plant was designed in 1963 to handle 0.025 mil m^3/day (6.5 mgd) municipal wastewater from a population of 35,000 and effluent from the sanitary facilities at the Dow Chemical Co. complex nearby. Every year since 1972, the actual average flow has exceeded the design average. For the year July 1975 through June 1976, the flow averaged 0.029 mil m^3/day (7.63 mgd). A flow equalization basin at the head of the plant evens out daily variations in flow, thus avoiding the need for a total plant capacity increase.

A high degree of treatment is provided by the wastewater system, which incorporates two-stage trickling filters, tertiary multi-media filtration, and phosphorus removal by chemical addition. The system provides approximately 90 percent reduction of both BOD and SS. Average influent BOD and SS concentrations are about 106 and 151 ppm, respectively. Industrial wastes comprise an insignificant part of the plant influent and create no wastewater or sludge treatment problems.

History

Historical plant modifications which have affected sludge production and characteristics are listed below:

May 1972 - Thermal sludge conditioning replaced polymer conditioning.

March 29, 1973 - Ferric chloride and polymer addition to primary influent for phosphorus removal started.

May 1, 1973 - Discontinued thermal conditioning and returned to polymer conditioning.

May 29, 1973 - Raised reactor temperature of thermal conditioning unit from 185° to 202°C (365°F to 395°F).

June 1973 - Tertiary mixed media filters added; backwash water recirculated to head of plant.

Chemical Addition for Phosphorus Removal

Liquid ferric chloride (40 percent $FeCl_3$) is added to the raw sewage between the raw sewage comminutor and the raw sewage pumps. Flow-proportional feed pumps are set to achieve a concentration of approximately 42 mg/ℓ $FeCl_3$. The turbulence created by the pumps aids in mixing the ferric chloride. Ferric phosphate begins to form, and the mixture enters a grit chamber where an anionic polymer supplied by Dow Chemical Co. is added for coagulation. The usual dosage of dry polymer is 0.17 mg/ℓ. The contact time between the ferric and polymer addition is 3 to 4 min. After polymer is added there is another 1.5 min or so contact time before the flow enters the primary clarifiers. Approximately 1.05 t (1.16 tons) per day dry $FeCl_3$ and 4.3 kg (9.5 lb) per day dry polymer are used. The plant's average phosphorus removal efficiency is approximately 80 percent, and phosphorus is generally reduced to a final effluent concentration of less than 1 mg/ℓ, meeting discharge permit requirements.

General Description of Wastewater Treatment Operations Affecting Sludge

Figure C-23 presents a general treatment plant flow diagram for Midland. It can be seen that sidestreams from the sludge treatment processes, backwash water from the tertiary filters, and waste sludge from the intermediate and final clarifiers all join the raw sewage before it enters the primary clarifiers. The clarifiers themselves are six rectangular tanks, with a total capacity of 0.0023 mil m^3 (610,500 gal). Each tank is 25.9 m (85 ft) long by 4.8 m (16 ft) wide by 3.05 m (10 ft) deep. The bottoms of the tanks are flat. Sludge is continuously scrapped along the bottom of each tank by a drag conveyor with seventeen flights to a hopper at one end. Each hopper is 4.88 m (16 ft) long, 1.22 m (4 ft) deep, 1.4 m (4 ft 6 in) wide at the top and 0.635 m (2 ft 1 in) wide at the bottom. Sludge is pumped from each tank once every 4 hr for 8 min, at 0.03 m^3/min (75 gal/min). In the past, sludge had been pumped for 15 min instead of 8. Decreasing the pumping time allowed a longer sludge detention time in the clarifier bottoms, which resulted in a sludge solids concentration of 7 percent TS instead of the former 6 percent TS. This increase in sludge solids concentration has had no effect on sludge handling operations except that less sludge volume is now pumped to the holding tanks and little or no supernatant is formed in the holding tanks and returned to the primaries.

It is clear that adding ferric chloride and polymer ahead of the primary clarifiers has affected primary clarifier performance and operation by increasing BOD, SS, and total phosphorus removal efficiency, and by increasing the amount of sludge solids generated. However, the amounts of these increases are difficult to determine. Other changes in plant operation took place about the same time that phosphorus removal was started. These changes affected clarifier efficiency and sludge production, so the effect of phosphorus removal alone is obscured. These plant modifications were the shortening of the sludge pumping time, affecting sludge solids concentration as already discussed, and the recycling of tertiary filter backwash to the head of the plant. The backwash water contains the solids removed in the tertiary filters. An unknown fraction of these solids is removed in the primary clarifiers, thus increasing sludge production by an unknown amount.

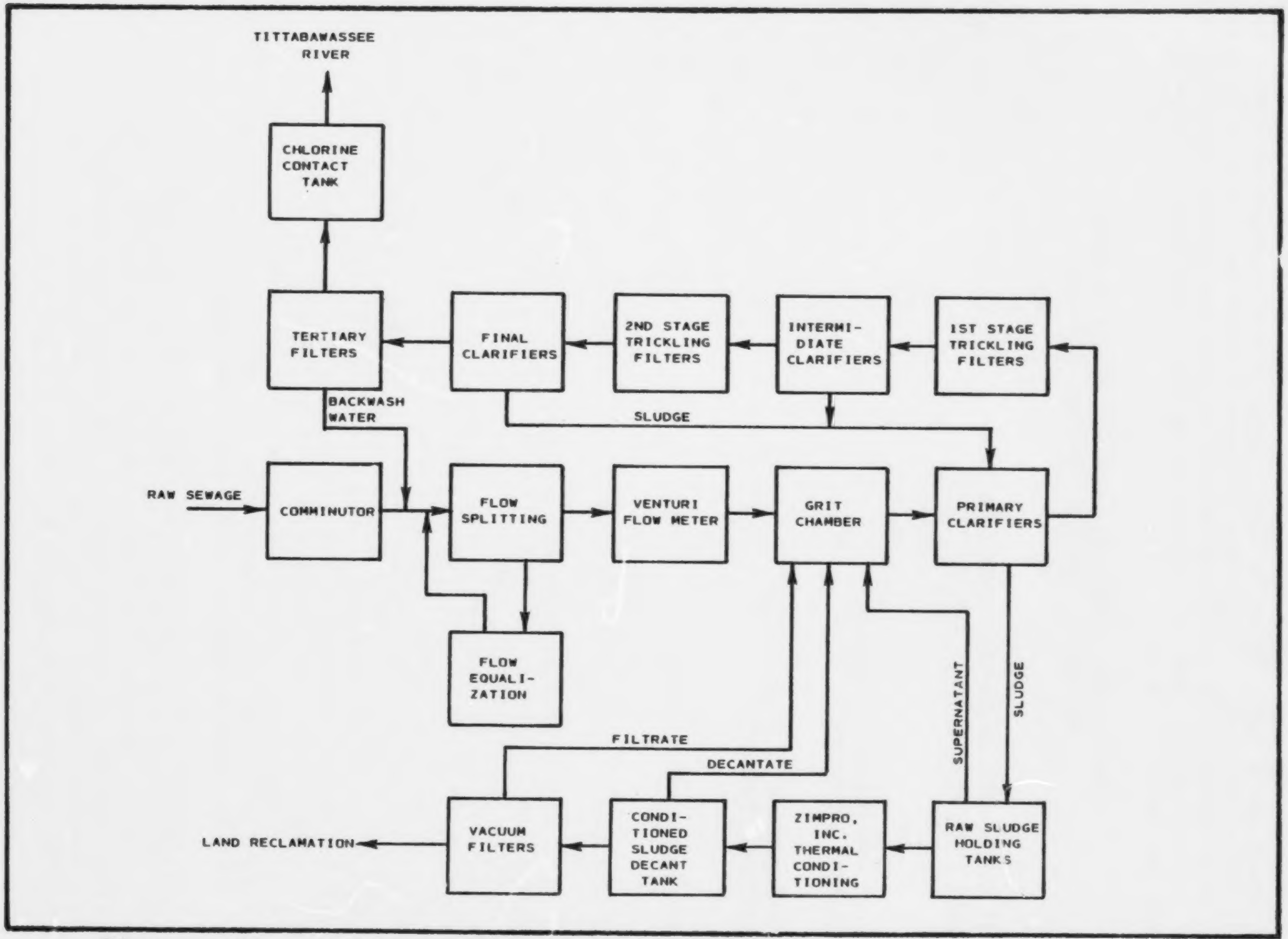

TITTABAWASSEE RIVER
CHLORINE CONTACT TANK
TERTIARY FILTERS
FINAL CLARIFIERS
2ND STAGE TRICKLING FILTERS
INTERMI-DIATE CLARIFIERS
1ST STAGE TRICKLING FILTERS
BACKWASH WATER
SLUDGE
RAW SEWAGE
COMMINUTOR
FLOW SPLITTING
VENTURI FLOW METER
GRIT CHAMBER
PRIMARY CLARIFIERS
FLOW EQUALI-ZATION
SUPERNATANT
SLUDGE
FILTRATE
DECANTATE
LAND RECLAMATION
VACUUM FILTERS
CONDI-TIONED SLUDGE DECANT TANK
ZIMPRO, INC. THERMAL CONDI-TIONING
RAW SLUDGE HOLDING TANKS

Figure C-23. Midland, Michigan, wastewater treatment plant flow diagram.

There was one month of operation between the commencement of phosphorus removal and the start-up of the tertiary filters. During that month only 19 mg/ℓ $FeCl_3$ were added instead of the present 40 to 50 mg/ℓ. Comparing primary clarifier efficiency during that month with the performance before phosphorus removal, we see increased BOD and SS removal with $FeCl_3$ addition. This is shown in Table C-29.

TABLE C-29. PHOSPHORUS REMOVAL IMPACTS ON PLANT BOD AND SS REMOVALS, MIDLAND, MICHIGAN

		No $FeCl_3$ Added	19 mg/ℓ $FeCl_3$
Total mil m^3 (MG) wastewater treated/mo:		0.94 (249.21)	0.95 (251.68)
percent removed by primary treatment -	BOD	26	48
	SS	32	67
Percent removed by first-stage biological treatment -	BOD	34	38
	SS	29	42
Percent removed by second-stage biological treatment -	BOD	51	38
	SS	14	37
Total plant percent removal -	BOD	62.4	80.2
	SS	63.9	89.0

The effluent from the six primary tanks flows to two first-stage stone-filled trickling filters, 24.4 m (80 ft) in diameter by 1.82 m (6 ft) deep. The effluent from these is collected in the underdrain system and flows to two intermediate clarifiers. Each clarifier is 18.3 m (60 ft) in diameter by 3.04 m (10 ft) deep. Trickling filter humus settles in the tanks and is continuously scraped to the center hoppers. Sludge is pumped from each hopper for 1 hr once every 8 hr. The sludge is returned to the primary clarifier influent. The effluent is pumped to the second stage trickling filters which are just like the first-stage filters. The final clarifiers are 21.9 m (72 ft) in diameter by 2.74 m (9 ft) deep. Sludge is removed from them as in the intermediate clarifiers and returned to the primary clarifier influent. The flow is then pumped to tertiary multi-media filters.

It appears that phosphorus removal has improved overall plant performance with regard to BOD and SS removal. Because of the increased removals in the primary clarifiers, the BOD and SS loadings on the secondary part of the plant are decreased, apparently enabling better trickling filter performance. This is borne out by the data in Table C-29.

Detailed Description of Sludge Treatment and Disposal Operations

Storage--

The two raw sludge holding tanks have served as thickeners in the past, but now serve only as holding tanks since thickening is occurring in the bottoms of the primary tanks. Each tank is square, 4.41 m (14.5 ft) on a side, and 6.09 m (20 ft) deep. Each has a capacity of 59.5 m^3 (15,710 gal). Sludge is pumped from these tanks to the thermal conditioner when it is in operation.

Conditioning--

Thermal conditioning is practiced with the Zimpro, Inc., low-pressure oxidation system. The principal difference between this system and the Porteous or Farrer processes is the addition of air to the reactor during the treatment operation. The system is really a low-pressure form of wet-air oxidation. Sludge is first ground to eliminate large particles. High pressure pumps then bring the sludge to a system pressure of 400 psig, before it is mixed with air from an air compressor. A heat exchanger transfers heat to the incoming sludge-air mixture from the outgoing sludge. The heated sludge then passes to a reactor, where steam injected from a boiler brings it to a temperature between 177°C and 204°C (350° and 400°F). The sludge leaves the reactor after 20 min, passes back through the heat exchanger, and enters two sludge thickening tanks. The combination of heat, air, and pressure in the reactor results in a cellular breakdown of the sludge solids, allowing a further gravity thickening to occur readily in the tanks.

Sludge is pumped to the thermal conditioner at 2.08 ℓ/sec (33 gal/min) when it is in operation. This is about 2.8 days/wk on the average, for 24 hr/day. The reactor temperature is presently maintained at 202°C (395°F). Approximately 14,000 m^3 (500,000 ft^3) of natural gas/mo are used to fuel the boiler.

Thermal sludge conditioning has several benefits. It stabilizes the sludge, enabling further handling without pathogens, odors, or putrefaction. It changes the cellular structure of the sludge, enabling thickening to a relatively high solids concentration and thus reducing the volume to be vacuum filtered. And it improves the dewatering characteristics of the sludge, increasing filter yield and cake solids concentration.

Thickening--

After thermal conditioning the sludge is thickened and stored in two tanks which are constructed like the two raw sludge holding tanks. Conditioned sludge is pumped into tank #1 until it is full. The decantate is then drawn off, and the thickened sludge is pumped over into tank #2. There is a screw mechanism in the center of tank #2 which is turned on whenever sludge is pumped to the vacuum filters. It mixes the denser sludge near the bottom of the tank with the less dense sludge near the top. The solids retention time in the thickener varies considerably between about 1.5 and 3 days because of intermittent vacuum filter operation. Sludge retention time and sludge blanket depth are not critical control variables for the operation of a conditioned sludge thickener as they often are for the operation of a raw sludge gravity thickener.

Dewatering--

Dewatering is accomplished by two Eimco rotary drum vacuum filters. Some machine operating parameters for these vacuum filters are given below:

Filter area: 18.6 m^2 (200 ft^2)
Filter media: polyethylene cloth, 11 x 12.6 threads/cm (28 x 32 threads/in)
Vacuum: 50.8 cm (20 in) Hg

The drum speed is set at 3 min per revolution, and the depth of drum submergence is 5.08 to 10.2 cm (2 to 4 in). Both filters are usually run at once. They are operated 3.7 days/wk for an average of 4 hr/day.

Chemical conditioning with anionic and cationic polymers has been used in the past. Approximately 0.150 kg (0.3 lb) of anionic polymer/t (ton) of dry solids and 2.50 to 4.00 kg (5 to 8 lb) of cationic polymer/t (ton) of dry solids were used. Thermal conditioning eliminated the need for chemical conditioners.

Ultimate Disposal--

Until about 1973, vacuum filter cake was buried in trenches at the city of Midland landfill site. A private contractor was hired by the plant to dig the trenches and bury the sludge. The filter cake at that time was unstabilized sludge. With thermal conditioning it became possible to use the sludge in a variety of other more useful ways. The sludge cake is still trucked out to the sanitary landfill site in two 6-wheel drive liftainer trucks belonging to the plant. Each truck carries a 4.59-m^3 (6-yd^3) load, and an average of 1.8 loads are hauled per day. A round trip is approximately 13 mi long and takes about 30 min. The sludge is stockpiled at the landfill site where the city of Midland can utilize it. The expenses involved in further handling are borne by the city of Midland using their General Fund.

Some of the sludge is used for land reclamation at the landfill site. After a section of landfill has been filled with refuse to the desired elevation, it is capped with clay soil.

Sludge is disced into this clay soil covering to enrich and condition it. The soil is then seeded with grass. In the future, when landfilling has ceased, the site will be grass covered and suitable for recreational use.

Some of the sludge is informally made into compost by the city's Department of Forestry and used as soil conditioner in their on-site ornamental tree nursery. Sludge or compost is disced into the soil, greatly enriching it and enhancing tree growth. Both sludge-leaf compost and sludge-sawdust compost are made in piles with a minimum of attention and labor.

Impact of Phosphorus Removal on Sludge Treatment and Disposal Operations

The sludge treatment flow diagram below (Figure C-24) shows as completely as possible the sludge and sidestream quantities handled at Midland. Approximately 75.7 m^3 (20,000 gal) of raw sludge containing 4,540 kg (10,000 lb) TS (7 percent TS) are pumped to the raw sludge holding tanks each day. Approximately 53.0 m^3 (14,000 gal) of thickener decantate are removed and the remaining 22.7 m^3 (6,000 gal) of concentrated (18 percent TS) sludge are vacuum filtered. The decantate contains only about 475 ppm SS, so we have estimated that only a few hundred pounds TS are removed in the decantate and that the rest are filtered. The vacuum filter filtrate is relatively high in suspended solids, containing 8,000 to 9,000 ppm. Several hundred pounds TS may be lost in the approximately 11.40 m^3/day (3,000 gal/day) of filtrate. Approximately 8.26 m^3 (10.8 yd^3) filter cake (50 percent TS) are formed each day. The dry weight of this filter cake is approximately 4,090 kg (9,000 lb).

The impacts of ferric chloride addition to the raw sewage on sludge treatment and disposal can be judged using plant data collected since 1970. The impacts upon sludge conditioning, thickening, and dewatering characteristics have varied with the conditioning method used, the thermal conditioner temperature, and the recycling of tertiary filter backwash water. Table C-30 consolidates available data indicating these relationships. The values presented are based on monthly averages as recorded in plant monthly reports.

Sludge Production and Sludge Solids Concentrations--

As discussed earlier, the impact of phosphorus removal on raw sludge quantity and solids concentration is obscured by several factors:

1. Modifications of sludge pumping procedures which increased the detention time of the sludge in the primary clarifiers;

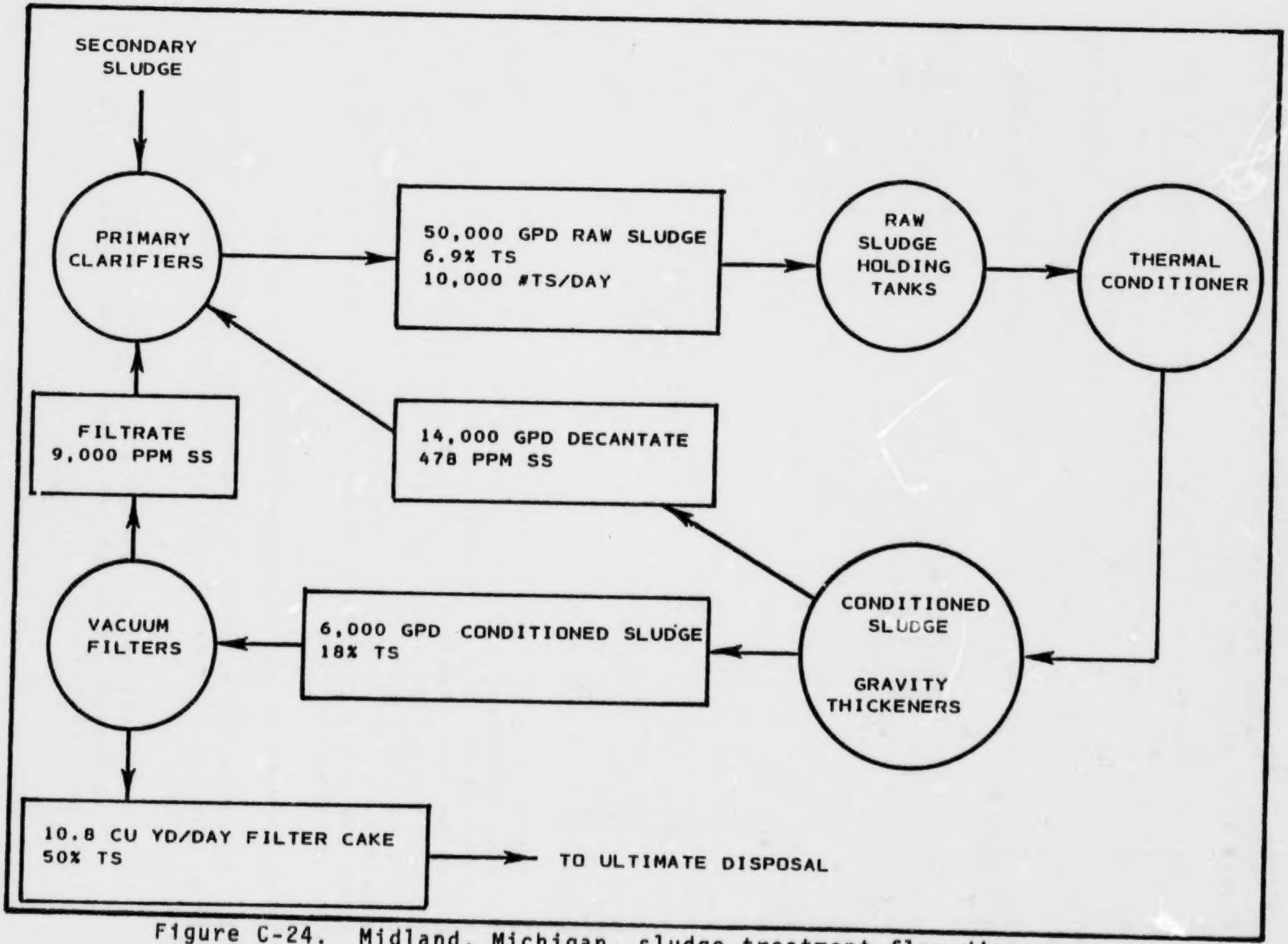

SECONDARY SLUDGE
PRIMARY CLARIFIERS
50,000 GPD RAW SLUDGE
6.9% TS
10,000 #TS/DAY
RAW SLUDGE HOLDING TANKS
THERMAL CONDITIONER
FILTRATE
9,000 PPM SS
14,000 GPD DECANTATE
478 PPM SS
VACUUM FILTERS
6,000 GPD CONDITIONED SLUDGE
18% TS
CONDITIONED SLUDGE
GRAVITY THICKENERS
10.8 CU YD/DAY FILTER CAKE
50% TS
TO ULTIMATE DISPOSAL

Figure C-24. Midland, Michigan, sludge treatment flow diagram.

TABLE C-30. IMPACTS OF PHOSPHORUS REMOVAL: SLUDGE CONDITIONING, THICKENING, AND DEWATERING CHARACTERISTICS--VARIABILITY WITH CONDITIONING METHOD, THERMAL CONDITIONER TEMPERATURE, AND THE RECYCLING OF TERTIARY FILTER BACKWASH WATER, MIDLAND, MICHIGAN

Variable	No FeCl$_3$ Polymer Cond.	No FeCl$_3$ Therm. Cond. 185°C (365°F)	19 mg/ℓ FeCl$_3$ Therm. Cond. 185°C	19 mg/ℓ FeCl$_3$ Polymer Cond.	25-60 mg/ℓ FeCl$_3$ Therm. Cond. 202°C (395°F) Tertiary Filters
1000 m^3 raw sludge/ mil m^3 sewage (1000 gal/MG)	4.15 (4.15)	2.74 (2.74)	2.92 (2.92)	N/A	1.98
% TS raw sludge	4.55	6.5	5.2	N/A	6.9
kg sludge TS/mil m^3 (lb TS/MG)	189,000 (1,575)	178,200 (1,485)	151,900 (1,266)	N/A	1,139 (13,6619.8)
Raw sludge VS (% of TS)	72	59	59	N/A	55.6
Polymer dosage-anionic kg/t dry solids (lb/ton)	0.15 (0.3)	0	0	N/A	0
Polymer dosage-cationic kg/t dry solids (lb/ton)	2.5 to 4.0 (5 to 8)	0	0	N/A	0
m^3 thickener decantate (1000 gal)	0	N/A	N/A	N/A	61 (16.1)
mg/ℓ SS thickener decantate	0	N/A	N/A	N/A	478
mg/ℓ BOD thickener decantate	0	N/A	N/A	N/A	6,000

(continued)

TABLE C-30 (continued)

Variable	No $FeCl_3$ Polymer Cond.	No $FeCl_3$ Therm. Cond. 185°C (365°F)	19 mg/ℓ $FeCl_3$ Therm. Cond. 185°C	19 mg/ℓ $FeCl_3$ Polymer Cond.	25-60 mg/ℓ $FeCl_3$ Therm. Cond. 202°C (395°F) Tertiary Filters
% TS sludge to filter	6	13	9	9.5	79
Filter yield-kg/m^2/hr (16/ft^2/hr)	23.9 (4.9)	39 (8)	20.2 (4.14)	14.7 (3.0)	78 (16)
mg/ℓ SS filtrate	647	4,057	5,682	3,469	8,288
Filtrate VSS (% of SS)	73.25	61.4	52	45.3	47
mg/ℓ BOD filtrate	2,724	7,795	N/A	2,950	6,769
% TS filter cake	25.5	48	44	39.3	54
Filter cake VS (% of TS)	72.25	57.9	56.4	47.8	46.4
m^3 cake hauled/t dry solids filtered (yd^3/ton)	4.03 (4.78)	3.00 (3.56)	3.27 (3.88)	3.46 (4.1)	2.66 (3.16)

2. Availability of only one month's data with $FeCl_3$ addition before the tertiary filters were placed in operation;

3. Tertiary filter operation, involving the recycling of backwash water containing the solids removed in the filters.

Sludge Thickening and Conditioning--

Raw sludge thickening to approximately 7.0 percent solids has been possible both with and without ferric addition. The thickening can occur in either the primary clarifiers or the raw sludge holding tanks. Thermal conditioning of the raw sludge changes the cellular structure of the sludge particles so that the sludge can be thickened even further prior to vacuum filtration. When there was no ferric addition for phosphorus removal, the thermally conditioned sludge thickened readily to 13 percent solids. With 19 mg/ℓ $FeCl_3$ addition, however, thickening to only 9 percent TS occurred. This adverse effect of ferric addition was overcome by raising the temperature of the thermal conditioner from 185°C to 202°C (365°F to 395°F). The sludge then thickened to 19 percent TS. The effect of ferric addition on the conditioned sludge thickener decantate is unknown. The decantate is currently low in SS (478 ppm), but high in BOD (6,000 ppm).

Sludge Dewatering--

Thermal conditioning has improved vacuum filter yield, raising it from 23.90 $kg/m^2/hr$ (4.9 $lb/ft^2/hr$) with polymer conditioning to 39.10 $kg/m^2/hr$ (8 $lb/ft^2/hr$) with thermal conditioning. The addition of 19 mg/ℓ $FeCl_3$ caused a decrease to only 20.2 $kg/m^2/hr$ (4.14 $lb/ft^2/hr$) for thermally conditioned sludge and 14.7 $kg/m^2/hr$ (3.0 $lb/ft^2/hr$) for chemically conditioned sludge. Raising the temperature of the thermal conditioner enabled a filter yield of 78.2 $kg/m^2/hr$ (16 $lb/ft^2/hr$).

The filtrate suspended solids concentration increased considerably with the switch to thermal conditioning. $FeCl_3$ addition apparently caused an even further increase. Tertiary filter operation may also have caused some increase because of the recycling of fine solids in the backwash water. The volatile fraction of the filtrate SS concentration appears to have decreased with the switch to thermal conditioning. This could be explained by the solubilization of volatile suspended solids during heat treatment and their removal in soluble form in the decantate. Concurrently, the BOD of the filtrate increased due to this solubilization. It appears that $FeCl_3$ addition further decreased the VSS fraction of the filtrate, and tertiary filter operation may have had a similar effect. It does not appear that the filtrate BOD was affected by $FeCl_3$ addition or tertiary filter operation.

The filter cake total solids concentration was greatly increased by thermal conditioning. Cake solids concentration was adversely affected by $FeCl_3$ addition, more so with polymer conditioning than with thermal conditioning. Raising the temperature of the thermal conditioner produced a dryer cake, somewhat dryer even than was achieved without $FeCl_3$ addition. The VS fraction of cake TS appears to have been decreased by thermal conditioning, with further decreases resulting from ferric chloride addition and from tertiary filter operation.

Sludge Cake Hauling--

As filter cake solids concentration increased, the volume of filter cake which must be hauled to the ultimate disposal site decreases. Table C-30 shows that the volume of cake hauled per ton of dry solids filtered was highest when polymer conditioning was used, and lowest when the thermal conditioner was operating at 202°C (395°F). With thermal conditioner operation at only 185°C (365°F), the adverse impact of ferric chloride addition resulted in more sludge cake volume to be hauled, and thus in more trips to the disposal site.

Experimental Addition of Alum to the Raw Sewage--

A full-scale test of alum addition to the raw sewage (80 mg/ℓ alum as Al_2O_3) was conducted at the plant during October of 1975. Enough alum was added to provide 80 percent phosphorus removal within the plant. Table C-31 presents the test results.

TABLE C-31. A COMPARISON OF THE CHARACTERISTICS OF THE SLUDGES PRODUCED WITH ALUM AND FERRIC CHLORIDE, MIDLAND, MICHIGAN

	Alum	$FeCl_3$
Raw sludge, % TS	6.5	7.6
Thermally conditioned sludge, % TS	16.0	22.5
Filter yield, $Kg/m^2/hr$ ($lb/ft^2/hr$)	23.5 (4.8)	76.7 (15.7)
Filter cake, % TS	41	56

Considerable problems were experienced in handling the sludge produced with alum. Separation of solids and supernatant in the thermally conditioned sludge thickener did not occur as rapidly as with ferric chloride. The TS concentrations of the thickened sludge was lower. The filter cake TS concentrations were only 41 percent compared to 56 percent with ferric chloride. Sludge with less than 50 percent TS is not suitable for the present method of disposal, indicating that alum sludge would require additional processing such as incineration. The filter rate was considerably slower with alum, it taking 3.3 times as long to filter a ton of alum sludge. This would mean adding one or possibly two full-time operators to the staff. Also, since the sludge pumped from the primary tanks was thinner, it was determined that the thermal conditioner would have to operate 17 percent longer to process it.

Sludge Treatment and Disposal Costs

In response to the adverse effects of phosphorus removal on sludge handling, the temperature of the thermal conditioner was raised. This restored satisfactory thickening and dewatering. Raising the thermal conditioner temperature from 185°C to 202°C (365°F to 395°F) meant an increase in the amount of natural gas required to fuel the boiler. The cost for natural gas increased by approximately $4.87/t ($4.34/ton) of dry solids as outlined in Table C-32. This is the only significant cost increase for sludge treatment and disposal which has been identified by this case study.

It has not been possible to establish the additional quantity of sludge solids generated by phosphorus removal, although it is certain that the quantity of sludge did increase. Without raising the temperature of the thermal conditioner, the additional sludge quantity would have been responsible for more vacuum filter running hours and more trips to the sludge cake disposal site. But at the higher thermal conditioner temperature, these trends were countered by a higher filter yield and cake solids content. It is therefore likely that the additional sludge quantity was handled without a significant increase in the costs of vacuum filter operation and sludge cake hauling. The plant presently runs the vacuum filters for about 94 hr/mo and makes about 79 trips/mo to the sludge cake disposal site.

Summary and Conclusions

Table C-30 has summarized the discussion of the variations in the performance of the sludge treatment processes. Thermal conditioning improved vacuum filter yield and cake dryness, but raised the BOD concentration of the thickener decantate and the filtrate and the SS concentration of the filtrate. Ferric chloride addition adversely affected filter yield, cake dryness, and filtrate SS concentration of the filtrate. Ferric chloride addition

TABLE C-32. ADDITIONAL COST FOR HIGH-TEMPERATURE THERMAL CONDITIONING, MIDLAND, MICHIGAN
(Assume sludge feed at 6.5% TS; natural gas cost of $0.089/m^3 ($2.50/1000 ft^3) in 1976)

	Before Phosphorus Removal 185°C (365°F) Cond. Temp.	After Phosphorus Removal 202°C (395°F) Cond. Temp.	Difference
1000 m^3 natural gas/mil m^3 sludge (1000 ft^3/MG)	4,365 (590)	7,841 (1,060)	3,477 (470)
$/mil m^3 sludge ($/MG)	$3,900 ($14.75)	$7,000 ($26.50)	$3,100 ($11.75)
$/t dry solids ($/ton)	$6.00 ($5.44)	$10.78 ($9.78)	$4.78 ($4.34)

adversely affected filter yield, cake dryness, and filtrate SS concentration. Even with the adverse impacts of ferric addition, however, filter yield and cake dryness were better with thermal conditioning than with polymer conditioning. Raising the temperature of the thermal conditioner overcame the adverse effects of ferric addition which were mentioned except for the increase in filtrate SS concentration. Filtrate SS concentration remained high, probably due to the recirculation of tertiary filter backwash water rather than due to ferric addition.

The decrease in vacuum filter cake volatile solids fraction (percent of TS) which occurred with thermal conditioning and further occurred with ferric addition would be an adverse impact for a plant practicing sludge cake incineration. Ferric chloride addition appears to have lowered the volatile fraction of cake solids from 58 down to 48 percent of TS.

A cost analysis has shown that the only significant cost increase for sludge treatment and disposal attributable to ferric addition was the cost of raising the thermal conditioner temperature to 202°C (395°F). The additional natural gas required to operate at the higher temperature had a cost of approximately \$4.78/t (\$4.34/ton) of dry solids (Table C-32).

Some general observations about this case study can be made. Apparently, the characteristics of the Midland sludge are such that it has thickened and dewatered readily both before and after phosphorus removal. It is likely that this is related to the facts that the plant uses trickling filters, has no significant industrial wastes, and is not overloaded. It has also been shown that ferric addition to the raw sewage produces a better sludge than alum addition (Table C-31).

In conclusion, thermal conditioning at Midland has been successful, but phosphorus removal has increased the energy requirements of the process at a time when energy is becoming more expensive.

CASE STUDY J: PORT HURON, MICHIGAN

Introduction

Port Huron is an example of an activated sludge plant practicing alum addition to the secondary part of the plant. Built from 1972 through 1974, it is representative of the newer wastewater treatment plants. Characteristic of these newer plants is the possession of adequate sludge handling capacity. At older plants, in contrast, problems related to lack of sludge handling capacity tend to dominate sludge handling considerations. Another characteristic of many newer plants is that phosphorus removal by chemical addition has been incorporated into their original design rather than added later. These plants have not handled sludge when no chemical sludge was being generated.

At Port Huron, sludge treatment consists of gravity thickening, thermal conditioning, polymer conditioning, centrifuge dewatering, and fluidized bed incineration. The system has been operated and evaluated both with and without inclusion of the thermal conditioning step. At plants where incineration seems to be the proper disposal method, thermal conditioning prior to dewatering has been cited as a way to reduce overall fuel costs. The combination of thermal conditioning and incineration has recently been chosen for the design of several new facilities. Centrifuges for dewatering are also finding application in these facilities. Investigation of the Port Huron facility gives us insight into the actual performance of an entire system using these techniques for handling phosphorus-laden chemical sludge.

Every plant has its peculiarities of design and influent wastewater characteristics. At Port Huron, the significant ones which affect sludge are related to the low flow rate and SS and BOD concentrations of the influent. Designed to provide secondary treatment for an average flow of 0.075 mil m^3/day (20 mgd), the plant treated an average flow of only 0.04 mil m^3/day (11.6 mgd) in 1976. Because of this discrepancy between plant design and actual flow, there is a long aeration basin retention time and a low food to microorganism (F/M) ratio. The activated sludge microorganisms are in a near-starvation phase and have a slow growth rate. This probably means that sludge generation is somewhat lower than typical for this type of plant.

Plant influent SS and BOD concentrations are quite low due to the fact that 20 percent of the wastewater entering the plant originates from industries, mainly of the metal manufacturing, coating, and plating types. Wastewater SS and BOD are usually about 86 and 58 ppm, respectively. Because of the low concentrations, these constituents are more efficiently removed by biological treatment and settling than by primary settling. The production of secondary relative to primary sludge is higher than would be the case if influent SS and BOD concentrations were greater. This is undesirable because the waste activated sludge tends to be harder to thicken than the primary sludge.

History

Historical plant modifications affecting sludge production and characterictics follow:

1972 - Initiated expansion of old primary plant and construction of secondary treatment facilities, phosphorus removal facilities, and solids handling system.

January 1975 - Commenced operation of new facilities, including phosphorus removal by chemical addition.

December 1976 - Discontinued use of heat treatment unit for sludge conditioning. Increased dosage of polymer conditioners used for dewatering.

Chemical Addition for Phosphorus Removal

Liquid alum (48.8 percent $Al_2(SO_4)_3 \cdot 14H_2O$) is added to the activated sludge mixed liquor as it flows over the effluent weirs of the aeration basins. Figure C-25 shows how wastewater enters and leaves the aeration tanks and the alum (A) and polymer (P) feed points. Alum is added at six points to the mixed liquor as it falls over the effluent weirs into six effluent channels leading to the final clarifiers. Alum is added at the farthest end of each weir from the final tanks, while polymer is added at the other end, 30.5 m (100 ft) closer to the finals. The alum is diluted as it is split into the six feed lines at a ratio of 5 to 1.

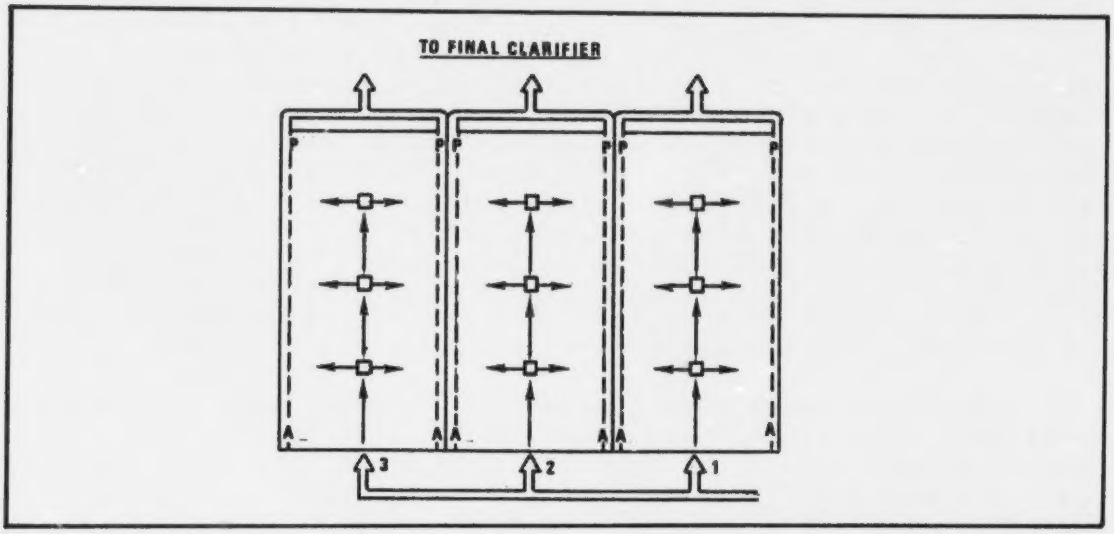


Figure C-25. Chemical feed points to aeration tanks, Port Huron, Michigan.

The alum dosage is adjusted according to plant phosphorus removal efficiency. Operators notice changes in removal efficiency and decide whether to increase or decrease the alum dosage. When a certain dosage is desired, a computer is signalled and the alum feed pump is automatically paced by the computer according to the wastewater flow to achieve that dosage. The average dosage has been approximately 52 ppm $Al_2(SO_4)_3 \cdot 14H_2O$ over the first 7 mo of 1977. Anionic polymer (Hercules 847) is used. The computer proportions the polymer feed pumping rate to the wastewater flow to achieve a desired dosage. The dosage is selected by plant personnel based on visual clarity of the final clarifiers and final effluent. The average dosage has been 0.275 ppm for January through July of 1977.

General Description of Wastewater Treatment Operations Affecting Sludge

Figure C-26 presents a general plant flow diagram. The plant has two aerated grit chambers complete with mechanical grit removal and mechanical oil and scum removal facilities. Eight primary clarifiers having a total capacity of 4,000 m^3 (138,400 ft^3) are designed to handle an average flow of 0.12 mil m^3/day (33 mgd) and a peak flow of 0.22 mil m^3/day (58 mgd). A flow-equalizing retention basin for primary effluent with 0.022 mil m^3 (5.7 MG) capacity is used. The clarifiers are rectangular. In each clarifier, sludge is scraped to a hopper at one end by a chain and flight mechanism. In the hopper, a screw mechanism scrapes sludge to a valve opening in the center. Sludge is collected from a pair of clarifiers at a time for 7 out of every 28 min. The flights and screws are turned on, a valve opens from two clarifiers and sludge is pumped from them at a combined rate of 0.984 m^3/min (260 gal/min) for the 7 min. Primary sludge is pumped at less than one percent solids for efficient degritting. Scum from the surfaces of the clarifiers enters a scum well which is discharged once a day to the thickener splitter box.

Secondary biological treatment is accomplished with conventional complete mix activated sludge. Three aeration basins providing a total capacity of 0.013 mil m^3 (3,350,000 gal) are mixed and aerated by diffused air. Their design detention time is 3 hrs, but the actual detention time in these basins averages about 6.9 hrs. The F/M ratio (figured using MLVSS rather than MLSS concentration) is maintained near 0.1 mg/ℓ. The return sludge rate averages 63 percent of flow. MLSS and MLVSS concentrations have averaged 5,250 and 3,000 ppm, respectively. Sludge density index averages around 1.7 percent.

Three square final clarifiers provide a total final settling capacity of 0.011 mil m^3 (2.9 MG). Sludge is sucked up hydraulically from the bottom of each clarifier by pickup pipes and flows by gravity to a manifold in the center of the clarifier bottom. Sludge is pumped out to return or waste from one clarifier at a time. The operator checks the depth

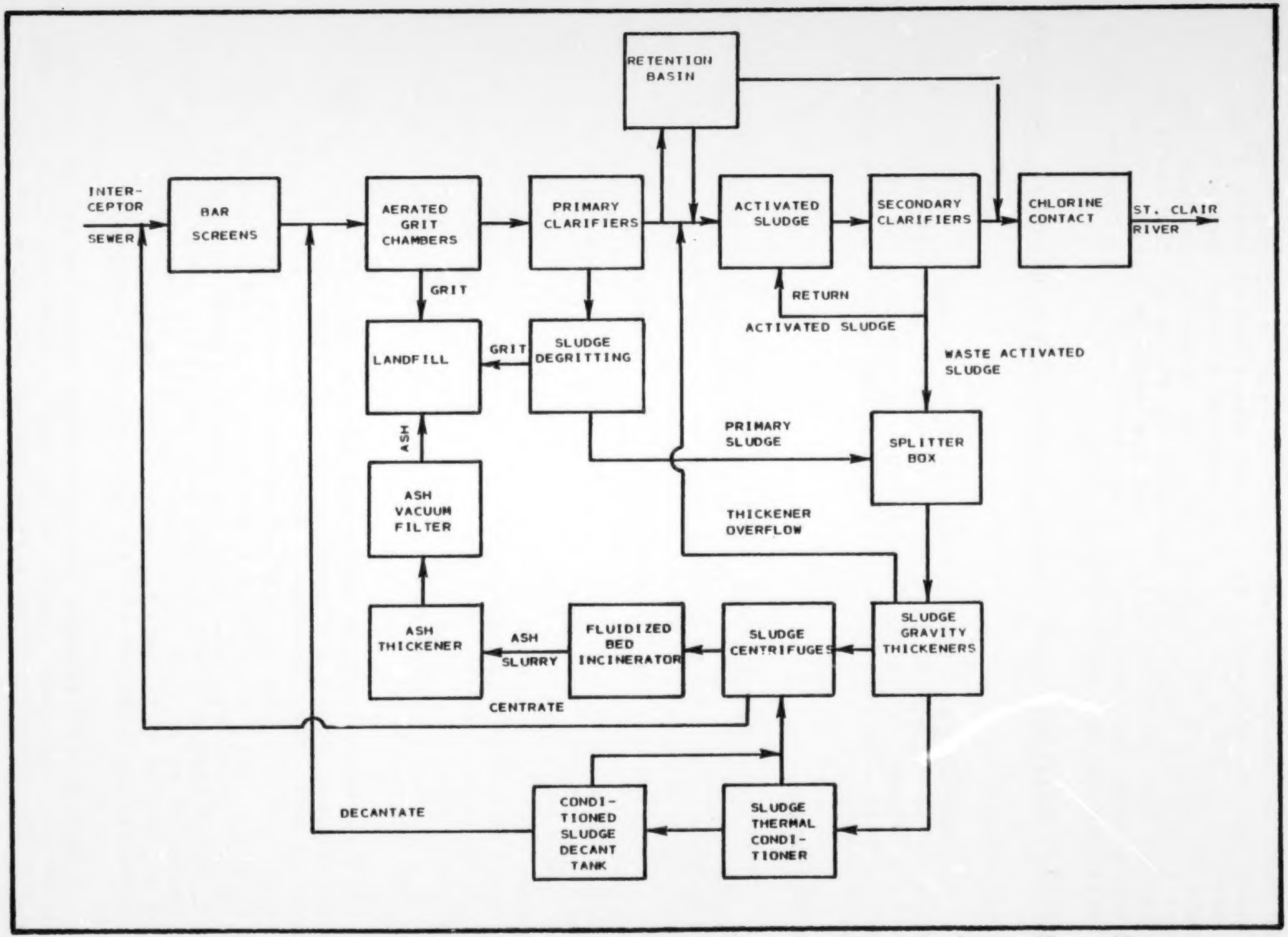

INTER-
CEPTOR
SEWER
BAR SCREENS
AERATED GRIT CHAMBERS
PRIMARY CLARIFIERS
RETENTION BASIN
ACTIVATED SLUDGE
SECONDARY CLARIFIERS
CHLORINE CONTACT
ST. CLAIR RIVER
GRIT
LANDFILL
GRIT
SLUDGE DEGRITTING
RETURN
ACTIVATED SLUDGE
WASTE ACTIVATED SLUDGE
PRIMARY SLUDGE
SPLITTER BOX
ASH
ASH VACUUM FILTER
THICKENER OVERFLOW
ASH THICKENER
ASH
SLURRY
FLUIDIZED BED INCINERATOR
SLUDGE CENTRIFUGES
SLUDGE GRAVITY THICKENERS
CENTRATE
DECANTATE
CONDI-TIONED SLUDGE DECANT TANK
SLUDGE THERMAL CONDI-TIONER

Figure C-26. Port Huron, Michigan, wastewater treatment plant flow diagram.

of the sludge blanket in each clarifier once a shift with the help of installed photoelectric cell detectors. He pumps sludge out as necessary to prevent the blanket depth from exceeding 0.914 m (3 ft). The average blanket depth is probably about 0.457 m (1.5 ft). The operator decides how much sludge to return and how much to waste on the basis of MLVSS in the aeration basins and desired F/M ratio. Scum is skimmed from the surfaces of the clarifiers into hoppers and discharged to the waste sludge lines.

Detailed Description of Sludge Treatment and Disposal Operations

Degritting--

A Dorr-Oliver "Dorr-Clone" cyclone sludge degritter is employed to remove 95 percent of 250 mesh grit from the primary sludge. Sludge degritting, in addition to the raw sewage grit removal, is necessary to limit wear on the sludge centrifuges.

Thickening--

Primary and secondary sludge, and scum from the primaries, secondaries, and aerated grit chamber flow to a Dorr-Oliver "Densludge" Type SD gravity thickener. Only one of the three thickeners present is used. A splitter box is present for dividing sludge flow between the three thickeners. A low cationic polymer is fed at the splitter box at a dosage of 3 ppm to aid solids capture and thickening. Each 16.8-m (55-ft) diameter thickener is of 738 m^3 (195,000 gal) capacity. The liquid sidewall is 3.35 m (11 ft) plus the cone at the bottom. It is equipped with scraper arms having vertical pickets for continuously stirring sludge in the thickener as it is scraped from the bottom to a hopper. Another scraper pushes sludge to a single draw-off point in the hopper from which sludge is pumped out to either the centrifuges or the thermal conditioner.

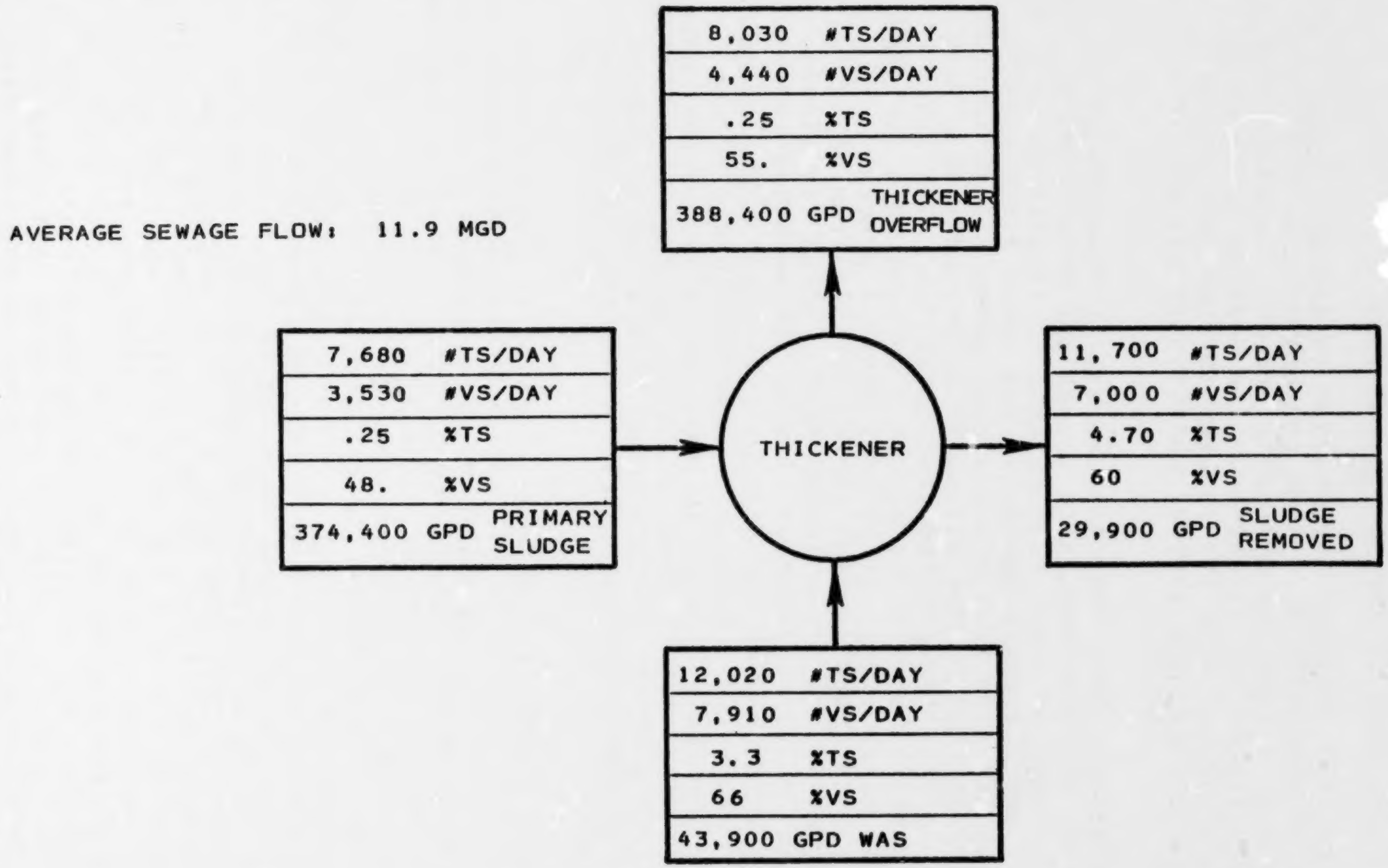
Thickener dilution water can be provided by three pumps which deliver final effluent to the splitter box. Dilution water is not used, however. Partial dilution is effected as a result of the low TS concentration at which primary sludge is pumped. Further dilution would be required to raise the thickener overflow rate into its design operating range of 16 to 32 $m^3/day/m^2$ (400 to 800 $gal/day/ft^2$). But the thickener is operated at a much lower overflow rate of approximately 7.13 $m^3/day/m^2$ (175 $gal/day/ft^2$). This is because its volume is effectively reduced by its high sludge blanket.

A mechanical skimmer collects scum from the surface of the thickener. The scum trough is so small, however, that once a day the thickener must be scraped manually. Scum adversely affects thickener operation by blocking the overflow weir and reducing weir overflow length. The scum baffle which is present is not deep enough to prevent this.

Figure C-27 presents information on average sludge flow rates, total and volatile solids concentrations, and masses in the form of a combined hydraulic and mass balance diagram constructed around the thickener. The diagram is based on monthly averages of data for January through July of 1977. During this period, the thermal conditioner was not in operation.

Figure C-27 shows that the average performance of the thickener is satisfactory, with a combined primary-waste activated sludge of about 0.56 percent TS thickening to 4.68 percent TS and the overflow containing about 2,480 mg/ℓ TS. But it does not show the inconsistency which exists in results from day-to-day due to variations in activated sludge wasting rates and intermittent incinerator operation. Thickened sludge concentration reaches almost 6 percent on a few good days, while on poor days it is only about 4 percent. The overflow solids concentration varies greatly from day-to-day between about 100 mg/ℓ and 10,000 mg/ℓ. These variations are related to the addition of alum to the aeration basins. Alum addition increases the mass of activated sludge which is generated and wasted. Waste-activated sludge has poorer thickening characteristics than primary sludge, so that when activated sludge wasting rates are high, thickening is poorer. This leads at times to low thickened sludge solids concentrations and high polymer dosage requirements for chemical conditioning.

Primary sludge is continually wasted to the thickener at 0.016 m^3/sec (260 gal/min). Activated sludge is normally wasted at 0.18 m^3/min (50 gal/min), but the rate varies between 0.096 and 0.36 m^3/min (25 and 100 gal/min). Wasting continues during all or part of the day, depending on how much needs to be wasted. Sludge is withdrawn from the thickener at an average rate of 0.18 m^3/min (52 gal/min) when the centrifuges and incinerator are operating. In 1977, they have operated 3 out of 4 wks/mo and for 2 to 5 days/wk (average 3.35 days/wk). A sludge blanket deeper than 1.2 (4 ft) is normally maintained in the thickener to get good thickening action. If the blanket is less than 0.61 m (4 ft), water can be drawn while pumping sludge, so this is avoided. The usual blanket depth is 1.5 to 2.1 m (5 to 7 ft). When the activated sludge wasting rate is high and the incinerator has not been running, a sludge blanket of up to 3.05 m (10 ft) builds up in the thickener, and overflow quality deteriorates. The thickener is in general slightly overloaded, and plans have been made to alleviate this by using one of the other two thickeners available. There is concern, however, about


AVERAGE SEWAGE FLOW: 11.9 MGD
8,030 #TS/DAY
4,440 #VS/DAY
.25 %TS
55. %VS
388,400 GPD THICKENER OVERFLOW
7,680 #TS/DAY
3,530 #VS/DAY
.25 %TS
48. %VS
374,400 GPD PRIMARY SLUDGE
THICKENER
11,700 #TS/DAY
7,000 #VS/DAY
4.70 %TS
60 %VS
29,900 GPD SLUDGE REMOVED
12,020 #TS/DAY
7,910 #VS/DAY
3.3 %TS
66 %VS
43,900 GPD WAS

NOTE. RATES ARE CALCULATED ON AN AVERAGE DAILY FLOW BASIS WHETHER FLOWS OCCUR EVERY DAY OR NOT.

Figure C-27. Port Huron, Michigan, gravity thickener hydraulic and mass balance.

increasing the solids detention time in the thickeners which, it was observed at the plant, results in poorer dewatering than can be achieved with "fresher" sludge.

Conditioning--

The two alternative methods of sludge conditioning which have been practiced are thermal conditioning with the Farrer system and chemical conditioning with polymers. Polymers were used in low concentrations even with thermal conditioning.

The Farrer system has two high-pressure feed pumps which force thickened sludge through two sludge disintegrators and to the Farrer heat exchanger. The preheated sludge flows to the Farrer reactor where steam, which has been purged of air, is injected. The sludge residues in the reactor for 20 to 30 min under 300 psi pressure at a temperature of 204°C (400°F). When sludge leaves the reactor, it flows through the heat exchanger to a decant tank. Decant tank overflow is returned to the head of the plant while sludge is withdrawn by the four centrifuge feed pumps. The system was designed to handle 20,600 kg/day (45,300 dry lb/day) of a 4.61 percent TS sludge feed. It was designed to produce a sludge capable of being centrifuged to 31 percent TS.

Polymer is added to the sludge in the lines which feed each of the four centrifuges at a point about 0.305 m (1 ft) before the sludge enters the hub of the scroll. Hercules 849 and 874 cationic polymers (1/2:1/2) are used. The average polymer dosage was 3.38 kg/t (6.75 lb/ton) of dry solids with thermal conditioning and 6.48 kg/t (12.94 lb/ton) without thermal conditioning.

With thermal conditioning, solids were processed (conditioned, dewatered, and burned) only about 25 percent of the time. The equipment was usually run 24 hr/day for 5 days and then kept down for 16 days. (Without thermal conditioning, solids were processed about 36 percent of the time.) Thermal conditioning was discontinued after November of 1976 due to extreme corrosion and erosion of the carbon steel making up the heat exchanger piping and connecting piping.

Thermally conditioned sludge from the decant tank averaged 8.6 percent TS. The decant tank overflow averaged abour 3,000 ppm SS with the volatile fraction averaging 45 percent. The pH of this sidestream varied between 6.0 and 7.2.

Dewatering--

Dewatering of thickened and chemically or thermally/chemically conditioned sludge is accomplished by centrifugation. The plant operates four horizontal fully-continuous solid bowl Dorr-Oliver "Merco-Bowl" centrifuges. Some machine operating parameters for these centrifuges are given below:

Largest inside bowl diameter: 41.9 cm (16.5 in)
Inside bowl length: 121.9 cm (48 in)
Bowl shape: conical-cylindrical
Scroll configuration: dual flight
Centrate disposition: pumped to plant influent

The machines are operated at a bowl speed of 3,000 rpm, scroll speed differential of 17 rpm and pool depth of 7.874 cm (3.1 in). When solids processing is underway, all four centrifuges are operated. Machine operating variables such as sludge feed rate and polymer dosage are monitored and controlled for each centrifuge separately by the operator. He makes adjustments which will enable at least 80 percent solids recovery. Centrate TS concentration is determined several times each day for each centrifuge.

Centrifuge and incinerator performance will be discussed together and related to type of sludge conditioning.

Incineration--

The plant has a Dorr-Oliver "F/S System" fluosolids incinerator. Centrifuge cake falls by gravity through chutes into sludge transfer screw conveyors. They convey it to the progressive cavity pumps which feed the reactor. The reactor cylinder is 4.57 m (15 ft) in diameter and contains a 1.52-m (5-ft) expanded bed of silica sand. A fluidizing air blower discharges into the reactor through the windbox. Hot gases and ash exit through the top of the reactor into a water seal expansion joint followed by a Venturi-type scrubber. They then enter a gas scrubber with a radial vane and water trays. The gases exit from the exhaust stack through a plume suppression burner. The ash slurry flows to ash pumps which deliver it to a 5.49-m (18-ft) diameter Dorr-Oliver ash thickener. Overflow from this thickener is recycled to the plant influent except for a portion which is used in the scrubber system. The ash is pumped to a vacuum filter (Dorr-Oliver) which is 1.220 m (4 ft) in diameter, has a 0.914 m (3 ft) face width, and is designed to handle 590 kg (1,300 lb)dry solids/hr. The filtrate is returned to the head of the plant. The dewatered ash is trucked 19.3 km (12 mi) to a landfill along with grit from the aerated grit chambers and sludge degritter.

Fluidizing air is normally discharged at 1.65 to 1.70 m^3/sec (3,500 to 3,600 scfm). The bed pressure differential is maintained at 1.19 to 1.22 m (47 to 48 in) of water. The normal reactor operating temperature is approximately 721°C (1,330°F). Although solids processing is intermittent, the incinerator temperature is always maintained above 663°C (1,225°F) over the next 30 hrs. It takes about 227 ℓ (60 gal) of No. 2 fuel oil to raise the temperature from 663°C to 871°C (1,225°F to 1,600°F).

The incinerator was designed to operate in conjunction with the thermal conditioner. Its design capacity is 1,040 kg (2,300 lb) of dry solids/hr for a 31 percent TS feed that is 58 percent volatile.

Performance and Operational Costs of Sludge Treatment and Disposal

Centrifuge and incinerator performance and operational costs have been greatly affected by the sludge conditioning method used. Table C-33 summarizes the direct effects of conditioning on the performance and operational costs of solids handling. Centrifuge cake TS concentration was improved by thermal conditioning and the pounds of solids which could be fed to the incinerator per hour was higher. This reduced incinerator fuel consumption. During thermal conditioning the burning rate covered a range of 420 to 1,320 kg TS/hr (929 to 2,900 lb TS/hr) which included the design capacity for thermally conditioned sludge of 1,040 kg/hr (2,300 lb/hr). Dorr-Oliver has estimated that the capacity is reduced to 590 kg/hr (1,300 lb/hr) for non-thermally conditioned sludge. Although total cake solids is higher, percent volatile is lower with thermal conditioning, due to the solubilization of volatile solids and removal in the decantate. This is a disadvantage as far as fuel consumption for incineration, but the disadvantage is outweighed by having less water in the cake to be evaporated. The major savings on utilities with thermal conditioning were for fuel oil and conditioning polymer. Major expenses were for natural gas and electricity and maintenance and repair of the thermal conditioner itself. Actual figures on maintenance and repair were not available based on the plant's experience because of the corrosion problem. Before normal operation could be achieved, the existing system would have to be rebuilt. The top five rows of the heat exchanger would have to be replaced with Schedule 160 inner tubes and Schedule 80 outer tubes, rather than the present Schedule 40 tubes. The total cost of labor for solids handling was the same with and without thermal conditioning. When thermal conditioning was discontinued, labor was shifted from that area to the incinerator.

The indirect effects of conditioning method on performance and operational costs of solids handling must also be considered. As mentioned, thermal conditioning solubilizes some of the volatile solids in the sludge which are removed in the decantate. The high BOD concentration of the decantate is a burden on the activated sludge treatment. The cost of

TABLE C-33. DIRECT EFFECTS OF CONDITIONING METHOD ON PERFORMANCE AND OPERATIONAL COSTS OF SOLIDS HANDLING, PORT HURON, MICHIGAN

Unit cost factors:

	(1976 costs)	
Fuel Oil	$ .1004/ℓ	$ 0.38/gal
Natural Gas	$87.07/1,000 m^3	$ 2.27/ft^3
Electricity	$ 0.03/kwh	$ 0.03/kwh
Centrifuge Polymer	$ 4.40/kg	$ 2.00/lb
Thickening Polymer	$ 2.97/kg	$ 1.35/lb
Gasoline	$ .185/ℓ	$ 0.70/gal

	With thermal conditioning	With chemical conditioning
kg TS/hr (lb TS/hr) to centrifuges and incinerator	777 (1,712)	556 (1,224)
kg VS/hr (lb VS/hr) to incinerator	350 (771)	342 (753)
Centrifuge cake % TS	27-32.6	18.8-21.0
Centrifuge cake % VS	45.05	61.6
Incinerator bed temp., ^{o}C (^{o}F)	738 (1,360)	732 (1,350)
kg conditioning polymer/t (lb/ton)	3.06 (6.75)	5.67 (12.5)
$/t ($/ton)	$14.88 ($13.50)	$27.51 ($24.96)
1,000 ft^3 gas used/ton	8.37	--
$/t ($/ton)	$20.94 ($19.00)	--
m^3 fuel oil used/t (gal/ton)	0.309 (74.1)	0.685 (164.1)
$/t ($/ton)	$31.04 ($28.16)	$68.75 ($62.36)
kwh electricity used/t (kwh/ton)		
thermal cond.	36.1 (32.7)	--
centrifuges	176 (159) - est.	176 (159) - est.
incinerator	374 (339) - est.	374 (339) - est.
thickener	9.92 (9) - est.	9.92 (9) - est.
$/t ($/ton)	$17.88 ($16.22)	$16.79 ($15.23)
Thermal cond. steam boiler water conditioning and cleaning chemicals est. $/t ($/ton)	$ 1.76 ($ 1.60)	--

(continued)

TABLE C-33 (continued)

	With thermal conditioning	With chemical conditioning
Odor control (therm. cond.) and sand replacement (incin.), est. $/t ($/ton)	$ 3.85 ($ 3.50)	$ 2.18 ($ 1.98)
Dewatered ash hauled, m^3/t (yd^3/ton)	0.961 (1.14)	0.961 (1.14)
est. $/t ($/ton) for gasoline	$ 0.08 ($ 0.08)	$ 0.08 ($ 0.08
kg thickening polymer/t (lb/ton)	0.926 (1.85)	0.926 (1.85)
$/t ($/ton)	$ 2.75 ($ 2.50)	$ 2.75 ($ 2.50)
Cost of labor,* $/t ($/ton) - est.		
1) Operational and minor equipment maint.		
thickener	$ 5.90 ($ 5.35)	$ 5.90 ($ 5.35)
thermal conditioner	$ 5.90 ($ 5.35)	--
centrifuges	$16.00 ($17.64)	$17.64 ($16.00)
incinerator	$16.00 ($17.64)	$23.54 ($21.35)
Total	$47.08 ($42.60)	$47.08 ($42.60)
2) General maintenance		
thickener	$ 3.53 ($ 3.20)	$ 3.53 ($ 3.20)
thermal conditioner	$ 1.76 ($ 1.60)	--
centrifuges	$ 5.30 ($ 4.81)	$ 5.30 ($ 4.81)
incinerator	$ 3.53 ($ 3.20)	$ 5.29 ($ 4.80)
Total	$14.12 ($12.81)	$14.12 ($12.81)
3) Equipment repair		
thickener	$ 0.49 ($ 0.44)	$ 0.49 ($ 0.44)
thermal conditioner	$ 0.49 ($ 0.44)	--
centrifuges	$ 1.95 ($ 1.77)	$ 1.95 ($ 1.77)
incinerator	$ 0.98 ($ 0.89)	$ 1.47 ($ 1.33)
Total	$ 3.91 ($ 3.54)	$ 3.91 ($ 3.54)
4) Ash and grit handling	$ 3.57 ($ 3.24)	$ 3.57 ($ 3.24)
Total maintenance and repair supplies for solids handling, $/t, ($/ton) - est.	$36.38 ($33.00)	$21.49 ($19.50)
Total supervision for solids handling, $/t ($/ton) - est.	$12.50 ($11.34)	$12.50 ($11.34)

(continued)

TABLE C-33 (continued)

	With thermal conditioning	With chemical conditioning
Total direct costs, \$/t (\$/ton)	\$210.79 (\$191.19)	\$220.78 (\$200.25)
thermal conditioner	\$ 33.62 (\$ 30.50)	--
centrifuges	\$ 45.04 (\$ 40.86)	\$ 57.68 (\$ 52.32)
incinerator	\$ 66.67 (\$ 60.49)	\$112.55 (\$102.09)
thickener	\$ 12.96 (\$ 11.76)	\$ 12.96 (\$ 11.76)

---

* Based on 1,090 t (1,200 tons) dry solids/yr in 1976. 1976 salaries and wages, including benefits.

increasing the power of the air blower to handle this BOD loading has been estimated at $9.52/t ($8.64/ton). The actual BOD concentration of the decantate is not available. Average centrate TS concentration was lower with thermal conditioning (966 vs 2,600 ppm) and volatile concentration was higher (63 vs 56.3 percent of TS).

Summary and Conclusions

Alum addition to the secondary rather than the primary part of the plant resulted in an increased mass of secondary rather than primary sludge. It is not known whether the solids concentration or volume of secondary sludge was affected by alum addition as well. Experience at the plant shows that the waste activated sludge is harder to thicken than the primary sludge. When the solids concentration of the thickened sludge is lower than average as a result of high activated sludge wasting rates, the cost of chemical conditioning prior to dewatering is increased. Solids concentration of the dewatered cake apparently does not vary in direct response to the variations in feed concentration, as long as adequate conditioning takes place and the centrifuge feed rate is properly adjusted. At the sludge loading and sludge removal rates in use, one 738-m^3 (195,000-gal) thickener is of insufficient capacity. This is evidenced by the high sludge blanket which sometimes builds up and causes deterioration of effluent quality. Rather than operate two thickeners, which might have deleterious effects on dewatering due to a long solids detention time, polymer was added to increase solids capture. It is recommended, however, that the design criteria for sizing thickeners be studied closely as problems in this area have been observed at several plants. Especially at plants where alum addition to the secondary stage is practiced, additional thickener capacity is needed.

The use of thermal conditioning resulted in savings of $50.30/t ($45.70/ton) of dry sludge when compared with the cost of chemical conditioning alone. The additional expenses (excluding equipment maintenance and repair supplies) resulting from thermal conditioning amounted to $35.00/t ($31.70/ton). Therefore, it appears that operation and maintenance of thermal conditioning at the plant are currently cost-effective, as long as the cost of equipment maintenance and repair supplies is less than $15.30/t ($13.92/ton).

Without thermal conditioning, the capacity of the incinerator is decreased from 1,040 kg/hr to 490 kg/hr (2,300 lb/hr to 1,300 lb/hr). It was assumed in the design of the incinerator that the plant would be producing 18,000 kg/day (39,700 lb/day) thickened sludge, but it actually produced only 5,310 kg/day (11,700 lb/day) due to the low wastewater flow. The incinerator capacity is therefore presently adequate at 590 kg/hr (1,300 lb/hr), but when the plant influent flow increases in the future, incinerator capacity will become limiting. It was estimated that three additional centrifuges and a 6.71-m (22-ft) diameter incinerator reactor rather than the present 4.57-m (15-ft) reactor would be necessary to handle 1,040 kg/hr (2,300 lb/hr) of sludge with chemical conditioning alone. Including this consideration in the analysis, operation and maintenance of chemical conditioning will not be cost-effective in the future unless additional capital costs are incurred.

Recent Developments

Since the original investigation of the Port Huron plant and the writing of this case study, modifications in thickener operation have been made with significant results. Two of the three identical gravity thickeners available at the plant are now in use rather than just one. Each of the two thickeners now receives approximately 12.6 ℓ/sec (200 gal/min) of dilution water plus 8.2 ℓ/sec (130 gal/min) of primary sludge and 1.6 ℓ/sec (25 gal/min) of waste activated sludge. They are operating at an overflow rate of approximately 22 m^3/day/m^2 (540 gpd/ft^2). The sludge blanket depth in each thickener is maintained between 1.2 and 1.5 m (4 and 5 ft).

In the past, where only one thickener was in use, the thickener was overloaded with sludge, although the surface overflow rate was low and no dilution water was used. The sludge blanket depth in each thickener averaged 1.5 to 2.1 m (5 to 7 ft). Polymer addition was necessary to aid solids capture and thickening. Presently, with two thickeners in operation, solids capture and thickening are reportedly slightly improved although no polymer is used. The savings in polymer cost amounts to $2.75/t ($2.50/ton) of dry solids. It is also reported that a reduction in the amount of phosphorus recycled to the head of the aeration basins in the thickener overflow has occurred. This may mean a slight decrease in the cost of phosphorus removal chemicals.

CASE STUDY K: PONTIAC, MICHIGAN

Introduction

The Sewage Treatment Division of the city of Pontiac, Michigan, operates two wastewater treatment plants a short distance from each other. The East Boulevard plant is a conventional activated sludge treatment plant with a design average flow of 32,600 m^3/day (8.6 mgd). The Auburn plant is also a conventional activated sludge treatment plant, with rapid sand filtration facilities for the tertiary treatment of both East Boulevard and Auburn secondary effluents. The Auburn Plant design average flow is 64,000 m^3/day (16.9 mgd), giving a combined design average flow for the two plants of 96,500 m^3/day (25.5 mgd). A flow diagram for the treatment process is shown in Figures C-28 and C-29. Wastewater influent to the Pontiac plants is delivered by combined storm and sanitary sewers. Approximately 40 percent of the influent consists of pretreated industrial process water (plating waste) from which heavy metals have been removed. This industrial waste is characterized by a relatively low concentration of BOD.

Two periods of plant operation have been selected for comparison purposes to determine the impacts of chemical additions for phosphorus removal. The period from January to June 1969 was selected as the period during which both primary and secondary addition of ferric chloride took place for phosphorus removal. Table C-34 presents the average wastewater characteristics and plant removal efficiencies of the two Pontiac plants combined.

History

In 1919, a secondary wastewater treatment plant, consisting of Imhoff tanks and trickling filters, was constructed at the East Boulevard site. In 1938, activated sludge aeration basins and an anaerobic digester were added. A primary settling tank and additional sludge treatment facilities were constructed in 1952. In 1962, a second plant with secondary treatment facilities was constructed at the Auburn site. The plant also contained anaerobic sludge digestion, vacuum filter dewatering, and incineration facilities. The Imhoff tanks and trickling filters at the East Boulevard plant were removed from service in 1962. In 1973, both plants were modified. A second primary settling tank and a flow equalizing retention basin were added at the East Boulevard site, and additional primary and secondary settling tanks and tertiary treatment facilities were added at Auburn.

Historical modifications to the plants affecting sludge are outlined in the following description:

- May 12-15, 1969 - Initial jar tests of phosphorus removal chemicals
- Aug.-Dec., 1969 - Full-scale testing of various chemical additions for phosphorus removal
- 1970 - Plants experimented with ferric chloride at the Auburn plant, with the use of various polymers
- 1971 - 1973 - Continuous addition of ferric chloride at the Auburn plant, with the use of various polymers

TABLE C-34. WASTEWATER CHARACTERISTICS AND REMOVAL EFFICIENCIES, PONTIAC, MICHIGAN

		Jan. - June '69	Jan. - June '77
Plant Flow m^3/day (mgd)		81,400 (21.5)	104,000 (27.6)
SS	Primary Influent (mg/ℓ)	184	105
	Primary Effluent (mg/ℓ)	182	66
	Plant Effluent (mg/ℓ)	23	3
	Primary Removal (%)	1	37
	Plant Removal (%)	87	97
VSS	Primary Influent (mg/ℓ)	100	62
	Primary Effluent (mg/ℓ)	106	36
	Plant Effluent (mg/ℓ)	13	2
	Primary Removal (%)	-6	42
	Plant Removal (%)	87	97
BOD	Primary Influent (mg/ℓ)	100	81
	Primary Effluent (mg/ℓ)	90	60
	Plant Effluent (mg/ℓ)	10	5
	Primary Removal (%)	10	26
	Plant Removal (%)	90	94
TP	Primary Influent (mg/ℓ)	-	3.8
	Primary Effluent (mg/ℓ)	-	1.9
	Plant Effluent (mg/ℓ)	-	0.2
	Primary Removal (%)	-	50
	Plant Removal (%)	-	95

- 1974 - Plant expansion begun (specific additions presented in general plant description summary); incinerator dismantled
- 1975 - 1976 - Various points of alum addition with and without polymer, plus secondary addition of ferric chloride used at the Auburn plant
- 1977 - Primary and secondary addition of ferric chloride at both plants; primary addition of anionic polymer at East Boulevard plant with some secondary addition of cationic polymer at Auburn plant
- March 1977 - Incinerator expansion completed and returned to service

Chemical Addition for Phosphorus Removal

The points of addition of the phosphorus removal chemicals were shown on Figure C-28. Liquid ferric chloride solution (approximately 41 percent $FeCl_3$) is added to the raw influent wastewater at a dosage of 8 to 12 mg/ℓ Fe^{+3}. Anionic polymer (Dru-Floc 2270) is mixed to achieve a 0.25 percent solution for addition at a dosage rate of 0.13 mg/ℓ to the raw sewage. Liquid cationic polymer is added at a dosage rate of 0.17 mg/ℓ to the aeration basin prior to secondary settling.

General Description of Wastewater Treatment Operations Affecting Sludge--

Table C-35 presents a description of wastewater treatment and sludge handling units at the Pontiac plants.

As Figure C-28 shows, sludge removed from the primary settling tanks is the only sludge sent directly to sludge processing by the anaerobic digester. Waste-activated sludge and plant sidestreams--sand filter backwash water, anaerobic digester supernatant, and vacuum filter filtrate--are all returned to the wastewater stream ahead of the primary settling tanks. Suspended solids are removed from the sludge and sidestreams in the primary tanks, from whence they are sent to the sludge processing units as part of the primary sludge.

Figure C-29 shows a sludge handling unit process flow diagram of the primary sludge removed from both the East Boulevard and Auburn plants. This processing consists of two-stage anaerobic digestion, followed by chemical conditioning, vacuum filtration, and incineration. Incinerator ash is disposed of either as a wet ash slurry in a lagoon on site, or as dry ash in a landfill.

TABLE C-35. GENERAL PLANT DESCRIPTION SUMMARY, PONTIAC, MICHIGAN

Aeration Tanks

(E) - 4 @ 54.6 m (179 ft) (1) x 6.1 m (20 ft) (w) x 3.4 m (11 ft) (SWD) per pass, two passes per tank; (A) - 4 @ 63.7 m (209 ft) (1) x 11.1 m (36.5 ft) (w) x 3.7 m (12 ft) (SWD); 2 @ 72.5 m (238 ft) (1) x 16.5 m (54 ft) (w) x 4.3 m (14.25 ft) (SWD).

Aeration Equipment

(E) - diffused air blowers - 294 m^3/min (10,500 ft^3/min), 8 surface. aerators @ 5 hp, 1,800 rpm; (A) - diffused air blowers - 280 m^3 (10,000 ft^3/min); 16 submerged turbine aerators at 10 hp, 37 rpm; 10 two-speed surface aerators at 40 hp, 68/45 rpm.

Secondary Settling Tanks

(E) - 4 @ 30.8 m (101 ft) (1) x 10.1 m (33 ft) (w) x 2.9 m (9.5 ft) (SWD); (A) - 2 @ 27.4 m (90 ft) x 2.6 m (8.5 ft) (SWD); 2 @ 27.4 m (90 ft) x 4 m (13 ft) (SWD); each tank equipped with: (E) - longitudinal sludge collector mechanismn with collector flights to sweep sludge on bottom of tank to sludge hopper at end of each tank, and rotating scum pipe to remove surface scum collected by flights on return travel; (A) - circular sludge collecting mechanisms with 10 sludge drawoff pipes each, which extend from V-ploughs on the bottom of the mechanism arms; and rotating, radial-arm scum collectors.

Anaerobic Digesters - Design Data

Plant	Type	Dia.	SWD	Cover	Mixing	Heating
E	Primary	22.9 m (75 ft)	8.2 m (27 ft)	Fixed	PFT Perth Mixer	1 external sludge heater
E	Secondary	22.9 m (75 ft)	8.2 m (27 ft)	Floating	None	Internal coils
A	Primary	21.3 m (70 ft)	7.6 m (25 ft)	Fixed	3 draft tubes	2 external sludge heaters
A	Secondary	21.3 m (70 ft)	7.0 m (23 ft)	Floating	None	None

Vacuum Filters

A - 2 Dorr-Oliver; 2 Eimco continuous cloth medium rotating drum filters; filter area - 34.8 m^2 (375 ft^2) each.

(continued)

TABLE C-35 (continued)

Filter Cake Converying and Storage System

Five conveyor sections - one in front of each pair of vacuum filters, one cross conveyor, one inclined to the top of the incinerator, and one across the top of the incinerator to the sludge filter cake hopper; storage hopper @ 61.m^3 (8 yd^3) capacity.

Incinerator

Multiple hearth furnace capacity - 5.4 t/hr (6 ton/hr) @ 25 percent TS prior to 1974, expanded to 7.3 t/hr (8 ton/hr) at 25 percent TS in 1977; number of hearths - 7; 10 auxiliary burners @ 2 per hearth in hearths No. 2, 3, 4, 5, and 6; minimum operating temperatures:

Hearth No.	Temperature °C.(°F.)
1	477.4 (900)
2	532.4 (1,000)
3	642.4 (1,200)
4	807.4 (1,500)
5	697.4 (1,300)
6	532.4 (1,000)
7	422.4 (800)

Center shaft and rabble arm drive @ 10 hp, 1.0 rpm; induced draft exhaust fan @ 100 hp, 358.4 m^3/min (12,800 ft^3/min); center shaft cooling air fan @ 10 hp, 106.4 m^3/min (3,800 ft^3/min).

Incinerator Exhaust Gas Scrubber

Flooded tray-type wet scrubber with quencher, scrubber, and Venturi; inlet gas temperature @ 422.4°C. (800°F.); exhaust gas temperature @ 37.4°C. (100°F.); 2 scrubber water supply pumps @ 2.3 m^3/min (610 gal/min).

Wet Ash Handling System

Ash hopper with mixer and vibrating screen; ash discharge pump 1.5 m^3/min (400 gal/min); agitator pump @ 0.23 m^3/min (60 gal/min).

(continued)

TABLE C-35 (continued)

Dry Ash Handling System

Two screw conveyors and one bucket elevator @ 5.6 m^3/hr (200 ft^3/hr); one rotary conditioner; one ash storage bin @ 78.4 m^3 (2,800 ft^3).

Notes:

E = East Boulevard Plant
A = Auburn Plant
DIA = Diameter
SWD = Side Water Depth

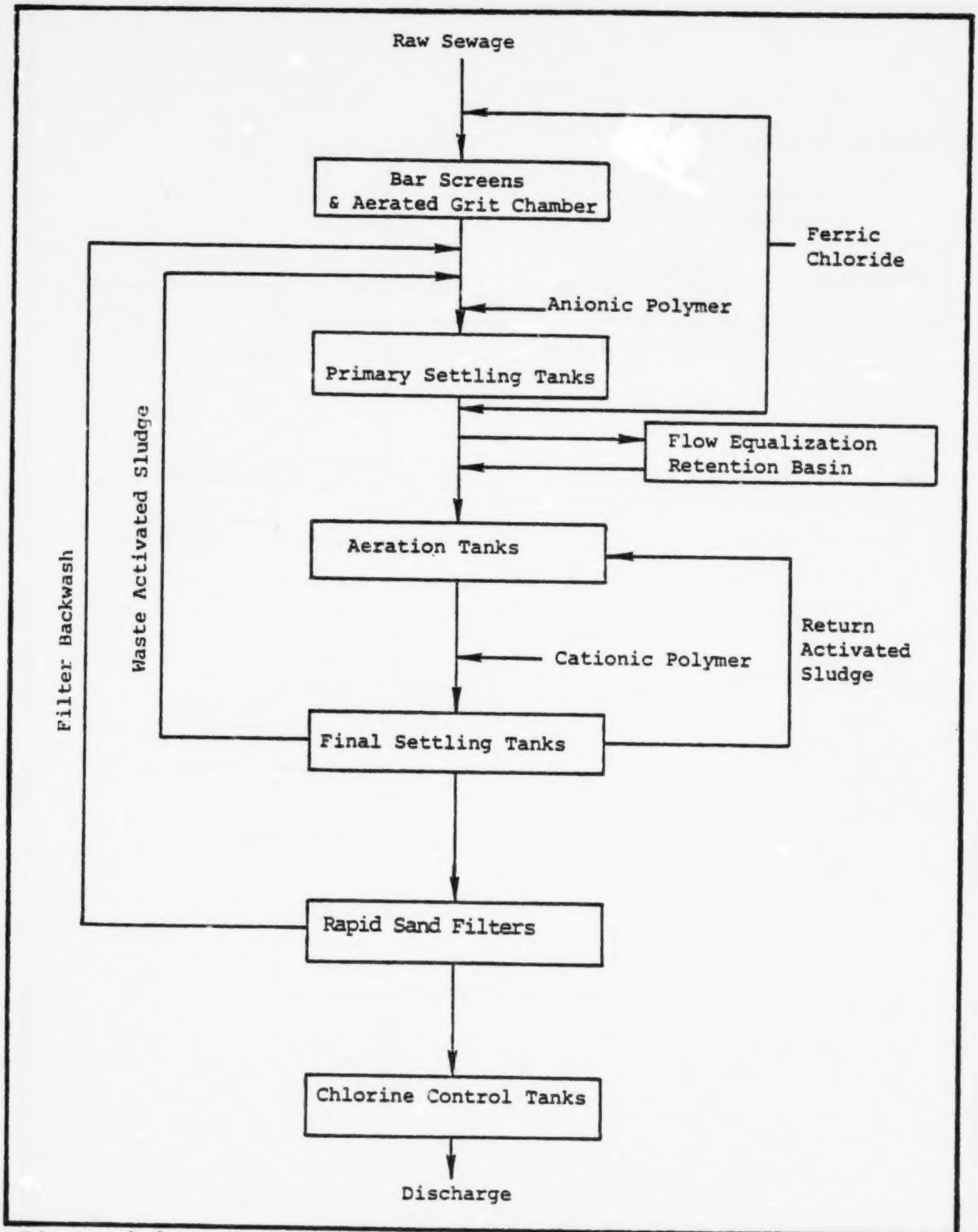

Raw Sewage
Bar Screens & Aerated Grit Chamber
Ferric Chloride
Anionic Polymer
Primary Settling Tanks
Flow Equalization Retention Basin
Aeration Tanks
Return Activated Sludge
Cationic Polymer
Final Settling Tanks
Waste Activated Sludge
Filter Backwash
Rapid Sand Filters
Chlorine Control Tanks
Discharge

Figure C-28. Wastewater treatment unit process flow diagram, Pontiac, Michigan.

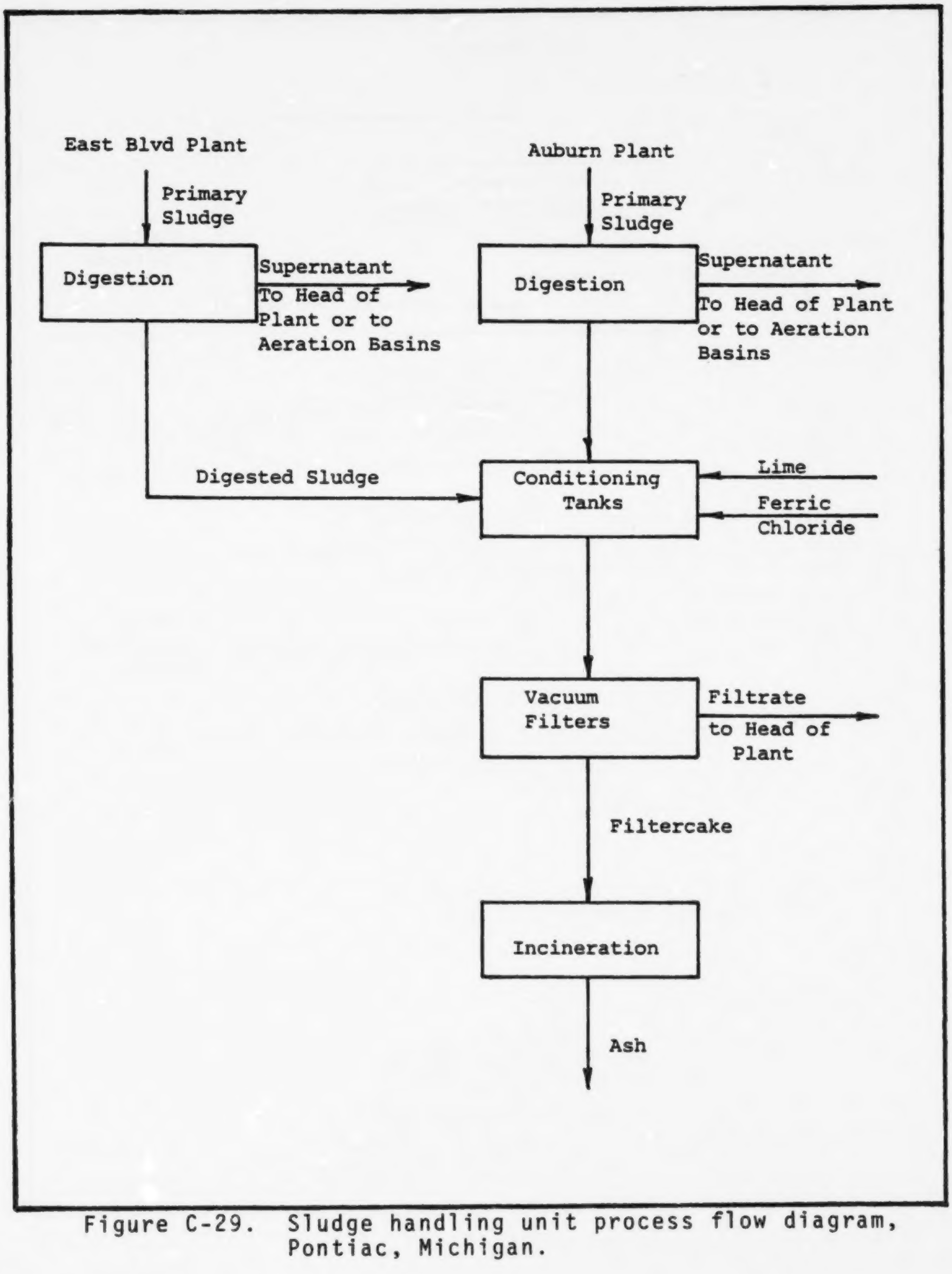

East Blvd Plant
Primary Sludge
Digestion
Supernatant
To Head of Plant or to Aeration Basins
Auburn Plant
Primary Sludge
Digestion
Supernatant
To Head of Plant or to Aeration Basins
Digested Sludge
Conditioning Tanks
Lime
Ferric Chloride
Vacuum Filters
Filtrate
to Head of Plant
Filtercake
Incineration
Ash

Figure C-29. Sludge handling unit process flow diagram, Pontiac, Michigan.

The lagooned ash is also sent to a landfill after drying. A materials balance of key wastewater treatment and sludge handling unit operations is superimposed on the plant flow diagram shown in Figure C-30. A brief description of each of the wastewater treatment unit processes which have significant impacts on sludge, follows:

Primary Clarifiers--

The primary clarifiers receive the raw influent wastewater and returned sidestreams, the most significant of which are characterized in Figure C-30. At the East Boulevard plant, sludge is removed on a continuous basis from the bottoms of the primary clarifiers to the sludge hopper. At the Auburn plant, sludge is removed during 3 out of every 4 hrs, 24 hrs/day from the bottom of the clarifiers to the wet well adjacent to each clarifier. Also, at the Auburn plant, sludge blankets regularly build up to a depth of 1.5 m (5 ft), although actual depths are not regularly monitored. Depths of the sludge blankets at the East Boulevard plant are also considerable, although not as great as those at the Auburn plant.

Sludge pumping from the sludge hoppers at East Boulevard and the wet wells at Auburn takes place during approximately 1 hr out of every 8 hr per clarifier. This sludge is then pumped to the primary digester at the respective plant.

Since the initiation of chemical addition for phosphorus removal, several significant changes in primary clarifier operations have been observed. Specifically, there have been significant improvements in primary clarifier removal efficiency of influent BOD and SS, as previously shown in Table C-34. Due to variations in plant influent characteristics, Table C-36 is presented to further highlight the changes in primary clarifier operations.

TABLE C-36. PRIMARY CLARIFIER WASTE STREAM CHARACTERISTICS, PONTIAC, MICHIGAN

	kg TS/kg SS (lb TS/lb SS) in Plant Influent	
Waste Stream	Jan. to June 1969	Jan. to June 1977
Digester supernatant	0.55 (1.21)	0.39 (0.86)
Waste activated sludge	0.65 (1.44)	<.65 (<1.44)*
Primary sludge (to digesters)	1.03 (2.26)	1.20 (2.65)

*Estimated due to reduced BOD loadings on secondary system.

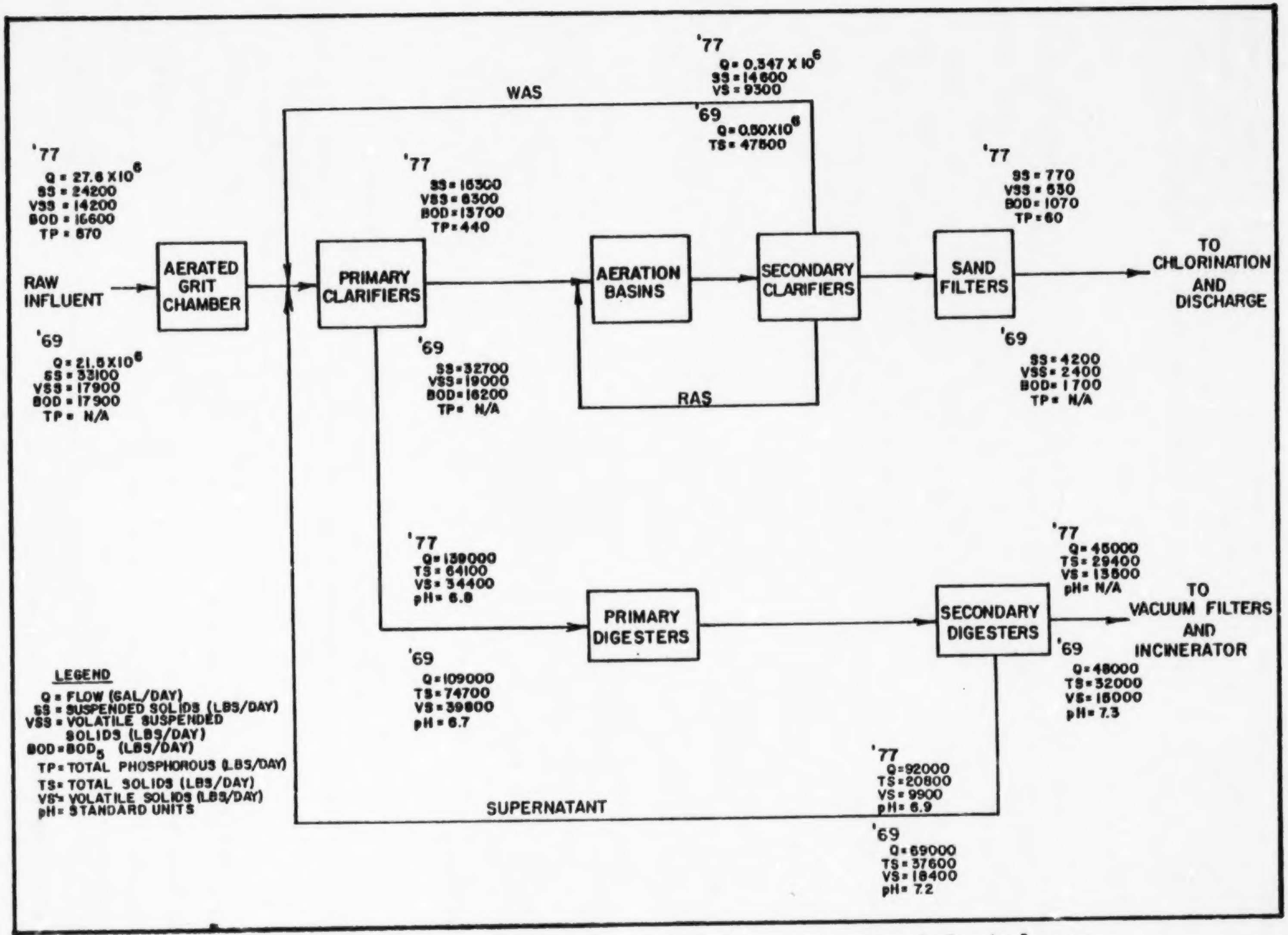

'77
Q = 27.6 X10^6
SS = 24200
VSS = 14200
BOD = 16600
TP = 870
RAW INFLUENT
'69
Q = 21.5X10^6
SS = 33100
VSS = 17900
BOD = 17900
TP = N/A
AERATED GRIT CHAMBER
WAS
PRIMARY CLARIFIERS
'77
SS = 16300
VSS = 8300
BOD = 13700
TP = 440
'69
SS = 32700
VSS = 19000
BOD = 16200
TP = N/A
AERATION BASINS
RAS
'77
Q = 0.347 X 10^6
SS = 14600
VS = 9300
'69
Q = 0.50X10^6
TS = 47800
SECONDARY CLARIFIERS
SAND FILTERS
'77
SS = 770
VSS = 630
BOD = 1070
TP = 60
'69
SS = 4200
VSS = 2400
BOD = 1700
TP = N/A
TO CHLORINATION AND DISCHARGE
'77
Q = 139000
TS = 64100
VS = 34400
pH = 6.8
'69
Q = 109000
TS = 74700
VS = 39800
pH = 6.7
PRIMARY DIGESTERS
SECONDARY DIGESTERS
'77
Q = 45000
TS = 29400
VS = 13600
pH = N/A
'69
Q = 48000
TS = 32000
VS = 16000
pH = 7.3
TO VACUUM FILTERS AND INCINERATOR
SUPERNATANT
'77
Q = 92000
TS = 20800
VS = 9900
pH = 6.9
'69
Q = 69000
TS = 37600
VS = 18400
pH = 7.2
LEGEND
Q = FLOW (GAL/DAY)
SS = SUSPENDED SOLIDS (LBS/DAY)
VSS = VOLATILE SUSPENDED SOLIDS (LBS/DAY)
BOD = BOD5 (LBS/DAY)
TP = TOTAL PHOSPHOROUS (LBS/DAY)
TS = TOTAL SOLIDS (LBS/DAY)
VS = VOLATILE SOLIDS (LBS/DAY)
pH = STANDARD UNITS

Figure C-30. Pontiac, Michigan, hydraulic and materials balance.

Significant decreases are apparent in the relative amounts of solids returned in the digester supernatant to the head of the primary clarifiers. A decrease in the quantities of solids contained in the waste activated sludge returned to the head of the primary clarifiers is also assured, since there were significant decreases in BOD loadings on the secondary treatment system. Overall, Table C-37 shows that sludge removed from the primary clarifiers contains significantly greater mass of solids, since chemical addition for phosphorus removal began in terms of the kilograms of TS per kilogram of SS in the plant influent.

Secondary Clarifiers--

Sludge is removed from the bottoms of the secondary clarifiers on a continuous basis at both the East Boulevard and Auburn plants. The majority of sludge withdrawn from the bottoms of these clarifiers is pumped as return activated sludge to the head of the aeration basins. A smaller portion is pumped as waste activated sludge to the head of the primary clarifiers to maintain the desired average mixed liquor suspended solids concentration. Since the initiation of chemical addition for phosphorus removal, there has been a considerable decrease in the volume of waste activated sludge per million gallons of plant influent.

Detailed Description of Sludge Treatment and Disposal Operations

Digesters--

Table C-37 describes the raw and digested sludge and supernatant characteristics before and after the initiation of chemical addition for phosphorus removal. During the period with dual point ferric addition, an increased volume of raw sludge at a lower average solids concentration was sent to the primary digesters. This led to a net increase in the mass of sludge TS fed to the digester per pound of SS in the plant influent. Most of the additional sludge TS fed to the digesters were volatile (fixed) solids.

Digester supernatant returned from the secondary digesters to the head of the primary clarifiers also followed the trend of increased volume at considerably lower solids concentrations. The net result was a significant decrease in the mass of supernatant TS per pound SS in the plant influent. These changes are indicative of improved digesting sludge settleability.

Lastly, the volume and solids concentrations of the digested sludge removed from the secondary digester were observed to have decreased. However, there were large increases in the mass of digested sludge TS per kilogram of SS in the plant influent.

TABLE C-37. RAW AND DIGESTED SLUDGE AND SUPERNATANT CHARACTERISTICS, PONTIAC, MICHIGAN

Raw Sludge	Jan.-June 1969	Jan.-June 1977
m^3/day (gpd)	413 (109,000)	526 (139,000)
% TS	8.2	5.5
% Volatile	53	54
pH	6.7	6.8
kg TS/kg SS influent to plant	2.26	2.65
kg VS/kg VSS influent to plant	2.22	2.42
kg FS/kg FSS influent to plant*	2.30	2.97
Supernatant		
m^3/day (gpd)	261 (69,000)	348 (92,000)
% TS	6.5	2.7
% Volatile	49	48
pH	7.2	6.9
kg TS/kg SS influent to plant	1.14	0.86
kg VS/kg VSS influent to plant	1.03	0.70
kg FS/kg FSS influent to plant	1.19	1.09
Digested Sludge		
m^3/day (gpd)	182 (48,000)	170 (45,000)
% TS	8.7	7.8
% Volatile	47	46
pH	7.3	N/A
kg TS/kg SS influent to plant	0.97	1.32
kg VS/kg VSS influent to plant	0.83	0.95
kg FS/kg FSS influent to plant	1.26	1.59

*FS = Nonvolatile (fixed) solids
FSS = Nonvolatile (fixed) suspended solids

Overall, the performance of the digesters improved significantly after the initiation of chemical addition for phosphorus removal. Digesting sludge settleability improved, as evidenced by major decreases in the mass of digester supernatant solids returned to the head of the primary clarifiers. Furthermore, the destruction of volatile solids influent to the digesters increased from 16 percent to 33 percent when the two periods are compared. A secondary impact of the increased volatile destruction was an increase in digester gas production from 1,960 up to 2,240 m^3/day (70,000 up to 80,000 ft^3/day).

Vacuum Filters--

Since the initiation of chemical addition for phosphorus removal, the vacuum filters have been operated 24 hr/day for an average of 13.8 days/mo, as opposed to 19.5 days/mo prior to the initiation of phosphorus removal. Table C-38 summarizes average vacuum filter performance for the two periods. During the period prior to the initiation of phosphorus removal, the vacuum filters were operated between 4 and 5 days/mo. During the period with dual point addition of ferric chloride to the wastewater, the vacuum filters were operated approximately 3 days/wk.

Since the initiation of chemical addition for phosphorus removal, there have been only minor changes in the dosage rates of conditioning chemicals added to the sludge prior to vacuum filtration. However, the filter yield rate has decreased from 19.8 to 17.8 g TS/hr (4.05 to 3.65 lb TS/ft^2/hr). In addition, the VS fraction of filter cake TS decreased from 33.5 to 30.5 percent of TS. Overall, however, there has been an increase in the capture of sludge solids in the filter cake. This has resulted in over a 40 percent decrease in the mass of filtrate solids returned to the head of the primary clarifiers.

Incinerator--

The multiple hearth incinerator is operated continuously during vacuum filter operations. Thus, the incinerator is operated 24 hr/day, 3 days/wk. Twelve additional hours are required to bring the furnace up to combustion temperatures, and 8 hours are required for cool-down. Table C-39 summarizes incinerator performance data for the two periods listed. It should be noted that the second period describes incinerator performance after the 1974 incinerator expansion was completed.

TABLE C-38. VACUUM FILTER PERFORMANCE, PONTIAC, MICHIGAN

	January - June 1969		January - June 1977	
	Average Per Day of Filter Operation	Average Per Day of Plant Operation	Average Per Day of Filter Operation	Average Per Day of Plant Operation
m^3 (gal) sludge fed	299 (79,000)	191 (50,500)	375 (99,000)	170 (45,000)
% TS of sludge	8.2	8.2	7.3	7.3
kg (lb) sludge solids fed	24,500 (54,000)	15,700 (34,500)	27,400 (60,300)	12,300 (27,000)
kg lime per kg sludge dry solids	0.143	0.143	0.156	0.156
kg ferric chloride/kg sludge dry solids	0.029	0.029	0.025	0.025
Filter yield (kg sludge TS/m^2 filter area/hr of operation (lb TS/ft^2/hr))	19.8 (4.05)	19.8 (4.05)	17.8 (3.65)	17.8 (3.65)
kg (lb) wet filter cake	74,500 (164,000)	47,700 (105,000)	94,900 (209,000)	43,100 (95,000)
% TS of filter cake	33.3	33.3	30.5	30.5
% VS of filter cake	37.7	37.7	39.0	39.0
kg (lb) filter cake dry solids	24,800 (54,600)	15,900 (35,000)	28,900 (63,700)	13,200 (29,000)
kg (lb) filter cake VS	9,310 (20,500)	5,990 (13,200)	11,300 (24,900)	5,130 (11,300)
kg (lb) filtrate total solids	3,950 (8,700)	2,540 (5,600)	3,670 (7,200)	1,500 (3,300)
% capture of sludge and conditioner solids	85	85	89	89

TABLE C-39. INCINERATOR PERFORMANCE,
PONTIAC, MICHIGAN

Incinerator Feed	Jan. to June 1969	April to June 1977
Hours of operation per month	368	348
kg (lb) wet cake/hr of operation	3,860 (8,500)	3,720 (8,200)
kg (lb) cake VS/hr of operation	499 (1,100)	454 (1,000)
m^3 (thousand ft^3) natural gas/mo	30.5 (1,090)	34.2 (1,220)
m^3 natural gas/kg wet cake (ft^3/lb wet cake)	0.02 (0.35)	0.03 (0.43)
m^3 natural gas/kg cake VS (ft^3/lb VS)	0.17 (2.70)	0.22 (3.55)

Since the initiation of chemical addition for phosphorus removal, there has been a decrease in the mass of wet cake fed to the incinerator per hour of operation. There has also been a decrease in the mass of volatile solids fed to the incinerator per hour of operation. As a result of these factors and incinerator capacity expansion, there has been an increase in the auxiliary fuel requirements per pound of wet cake and per pound of cake volatile solids fed to the incinerator.

Summary and Conclusions

There have been many changes in the Pontiac plant's wastewater and sludge handling operations as a result of chemical additions for phosphorus removal, changes in plant influent characteristics, and plant equipment additions and modifications. These impacts are highlighted below, for each of the unit operations listed:

Primary Clarifiers:

- Improved SS and BOD removal efficiencies
- Decreased solids loading on the primary clarifiers from the return of digester supernatant, waste-activated sludge, and vacuum filter filtrate
- Increased mass of primary sludge TS generated per kilogram SS influent to plant; additional sludge solids mainly nonvolatile.

Digesters:

- Increased volume of sludge pumped to digesters at lower average solids concentration, with net effect being an increase in the mass of primary sludge TS fed to the digesters per kilogram of SS in the plant influent
- Increased volume of supernatant, but at a considerably lower solids concentration, indicative of better sludge settleability within the digesters
- Increased volatile destruction occurring within the digesters
- Increased digester gas production.

Vacuum Filters:

- Decreased filter yield
- Decreased filter cake total solids and volatile solids concentrations
- Increased capture of sludge solids in the cake
- Decreased solids in filtrate returned to head of primary clarifiers.

Incinerator:

- Decreased feed rates in terms of kilograms wet cake and kilograms volatile solids per hour of operation
- Increased fuel consumption per hour of operation, per kilogram wet cake feed and per kilogram volatile solids fed.